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FLUIDIZED BED COMBUSTION PROCESS EVALUATION

PHASE I. RESIDUAL OIL GASIFICATION/DESULFURIZATION DEMONSTRATION AT ATMOSPHERIC PRESSURE. VOLUME I. SUMMARY

WESTINGHOUSE RESEARCH LABORATORIES

PREPARED FOR
NATIONAL ENVIRONMENTAL RESEARCH CENTER

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This volume of the report summarizes results of an evaluation of the atmospheric-pressure fluidized bed residual oil gasification/desulfurization process, a process referred to by its inventor (Esso Research Centre, Abingdon, England) as the chemically active fluidized bed (CAFB) process. The CAFB produces a clean, low heating value fuel gas for firing in a conventional boiler. The integrated process, previously operated successfully in a 750 kW pilot plant unit, has demonstrated the ability to meet environmental emission standards for sulfur oxides, nitrogen oxides, and particulates. Work carried out under this contract was directed toward completion of a preliminary design and cost estimate for a 50 MW demonstration plant and a 200 MW plant design and cost estimate. Several process and design options are evaluated. Process flow diagrams, energy and material balances, equipment specifications, vessel drawings, equipment arrangement drawings, a site plan, an electrical one-line drawing, and utility requirements are presented for the recommended process concept. Plant performance, environmental impact, and functional operating conditions are presented and development requirements identified. Capital and operating costs are presented for the 50 MW demonstration plant and for commercial plants with capacities from 50 to 500 MW. Limestone sorbent support data is given.

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Air Pollution	Boilers	Air Pollution Control	13B	13A
Combustion	Nitrogen Oxides	Stationary Sources	21B	07B
Fluid Bed Processing	Limestone	CAFB Process	13H, 07A	
Residual Oils		Particulates	21D	
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Desulfurization				

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FLUIDIZED BED COMBUSTION PROCESS EVALUATION

(PHASE I - RESIDUAL OIL GASIFICATION/DESULFURIZATION DEMONSTRATION AT ATMOSPHERIC PRESSURE)

VOLUME I - SUMMARY

by

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March 1975

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ABSTRACT

Westinghouse has carried out plant designs, cost estimates, process evaluation, and experimental work on an atmospheric-pressure fluidized bed residual oil gasification process for power generation. This volume is a summary of the work which is the first phase of a demonstration plant program. The process, conceived by Esso Research Centre, Abingdon, England, as the chemically active fluidized bed (CAFB) process, produces a clean, low heating-value fuel gas for firing in a conventional boiler. The integrated process, previously operated successfully in a 750 kw_e pilot plant unit, has demonstrated the ability to meet environmental emission standards for sulfur oxides, nitrogen oxides, and particulates.

Work carried out under this contract was directed toward a commercial demonstration of the process. A preliminary design and cost estimate for a 50 MW_e atmospheric-pressure fluidized bed residual oil gasification/desulfurization demonstration plant was completed for the 50 MW_e unit No.12 at the Manchester Street Station, Narragansett Electric Co., Providence, RI. The design and cost estimate provide sufficient detail to proceed with the detailed design and construction. Boiler performance with hot, low heating-value gas is assessed. Experimental programs were carried out to select limestone sorbents for the plant and to provide design and operating criteria on spent sorbent processing. The environmental impact of the process is assessed and areas which require further development identified. Market projections are presented for the process. A 200 MW_e commercial plant design and cost estimate were prepared. Capital and operating costs are presented for commercial plants with capacities from 50 to 500 MW_e. Detailed design and construction of the 50 MW_e demonstration plant utilizing low-grade, high metal content fuels is recommended if completion of the critical experimental tests identified and review of fuel projections demonstrate process operability and fuel availability.

PREFACE

The Office of Research and Development (ORD) of the United States Environmental Protection Agency (EPA) has organized and is sponsoring a fluidized bed fuel processing program. Its purpose is to develop and demonstrate new methods for utilizing fossil fuels to produce electrical energy from utility power plants which meet environmental standards. These methods should:

- Meet environmental goals for sulfur dioxide (SO_2), nitrogen oxide (NO_x), ash, smoke emissions, trace element emissions, and wastes
- Utilize fuel resources efficiently
- Compete economically with alternative means for meeting environmental goals.

Westinghouse Research Laboratories, under contract to ORD, is carrying out a program to evaluate, develop, and demonstrate fluidized bed gasification/desulfurization of residual fuel oil for power generation. This two-volume report describes work performed from May 1973 to December 1974 under contract 68-02-0605. The work carried out during this period is based on tasks set forth by EPA which were completed by Westinghouse under previous contracts. The results from these prior tasks on fluidized bed gasification of residual oil were published in a three-volume report, "Evaluation of the Fluidized Bed Combustion Process," in November 1971 under contract No. CPA 70-9[1] and as part of a four-volume report, Volume IV "Fluidized Bed Oil Gasification/Desulfurization," in December 1973 under contract 68-02-0217.[2] The previous work on residual oil gasification included:

- Conceptual design and cost estimate of atmospheric-pressure fluidized bed residual oil gasification/desulfurization system for utility power generation

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- Assessment of the effectiveness and economics of an atmospheric-pressure fluidized bed residual oil gasification/desulfurization system
- Identification of a project team to demonstrate fluidized bed residual oil gasification/desulfurization for power generation
- Identification of a boiler unit to carry out the demonstration plant program
- Evaluation of pressurized residual oil gasification/desulfurization for combined cycle power generation
- Provision of technical consultation and assistance on the fluidized bed fuel processing program.

Tasks carried out under contract 68-02-0605, which are reported in this two-volume report, have included:

- Preparation of a preliminary design and cost estimate for the 50 MW fluidized bed residual oil gasification/desulfurization demonstration plant. The design is for the 50 MW unit No. 12 at the Manchester Street Station, Narragansett Electric Company, Providence, Rhode Island. The design and cost estimate provide sufficient detail to proceed with the detailed design and construction of the demonstration plant.
- Evaluation of the demonstration plant boiler performance. The boiler performance with the hot, low heating-value gas is assessed for the demonstration plant. Modifications and associated costs to maintain the performance are identified.
- Evaluation and selection of the fluidized bed gasification process and design options. Alternative design and operating parameters are identified and evaluated. Recommendations are made for the demonstration plant and for commercial plants.
- Development of process data. Experimental tests were carried out to identify candidate limestone sorbents; to obtain data on sulfur removal, sorbent regeneration, and spent stone processing; and to assess the environmental impact of spent stone.

- Determine market for fluidized bed residual oil gasification/desulfurization. Factors are identified which will determine the market potential. Market projections are made based on a review of these factors.
- Prepare commercial plant design and cost estimate. A 200 MW commercial plant design is prepared on the basis of the demonstration plant design work. A factored cost estimate is made on the basis of the demonstration plant estimate.
- Prepare assessment of environmental impact. Solid, liquid, and gaseous emissions are identified and assessed to determine their environmental impact. Resource utilization is also reviewed.
- Identify Development requirements. Areas which require further development are identified and programs projected to obtain the necessary information to carry out the program.

This volume (Volume I) contains a summary of the work performed and program recommendations. Volume II contains appendices which provide supplemental information and detailed back-up to support the summary and recommendations.

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The results, conclusions, and recommendations presented in this volume represent the combined work and thought of many persons at Westinghouse and the Office of Research and Development (ORD), and personnel at New England Electric System (NEES). Other ORD contractors have freely shared with us their ideas and results of their research and development effort. Westinghouse subcontracted Stone and Webster Engineering Corporation (SWEC) to prepare the preliminary design and cost estimate.

In particular, we want here to express our high regard for and acknowledge the contribution of personnel at Westinghouse Research Laboratories, New England Power Service Company, Esso Research Centre (England), and the personnel at ORD who have directed the overall fluidized bed residual oil gasification program and who have defined, monitored, and supported the efforts of Westinghouse and others on the program. Mr. P. P. Turner, Chief of the Advanced Process Section, has served as project officer on our work. Numerous enlightening and helpful discussions have been held with Mr. Turner: with section members S. L. Rakes and D. Bruce Henschel; and with R. P. Hangebrauck, Chief of the Demonstration Projects Branch. Mr. S. K. Batra and Mr. R. Gendreau of New England Power Service Company contributed to the demonstration plant design, provided data on the Manchester Street Station, and participated in process evaluation and selection. Dr. J. W. T. Craig of Esso Research Centre (England) participated in the preliminary design study from August 1973 through February 1974. Dr. Craig was effective in integrating the Esso pilot-scale experience and results into the demonstration plant design. He also participated in the investigations and evaluation of spent stone processing systems. Mr. John P. Malloy of SWEC served as project manager for preparation of the preliminary design and cost estimate. Mr. Harry

Robinson, SWEC, was the lead process engineer. Allied Chemical Corporation provided technical and economic information on sulfur recovery. Mr. W. D. Hunter of Allied coordinated the efford. Many persons at the Westinghouse Research Laboratories have contributed to the program. Dr. T. K. Gupta and Dr. M. Gunasekaran are initiating work to identify and develop market applications for the spent stone. Mr. R. Brinza and Mr. W. F. Kittle assisted in the collection of data.

I. SUMMARY AND RECOMMENDATIONS

The atmospheric-pressure fluidized bed residual oil gasification/desulfurization process (CAFB)[*] produces a clean, low heating-value fuel gas for firing in a conventional boiler. The integrated process has previously operated successfully in a 750 kw[†] pilot plant unit, and the process has demonstrated the ability to meet environmental emission standards for sulfur oxides, nitrogen oxides, and particulates.

Work carried out under this contract was directed toward the completion of a preliminary design and cost estimate for a 50 MW demonstration plant and a 200 MW plant design and cost estimate. Several process and design options are evaluated. Process flow diagrams, energy and material balances, equipment specifications, vessel drawings, equipment arrangement drawings, site plan, an electrical one-line drawing, and utility requirements are presented for the recommended process concept. Plant performance, environmental impact, and functional operating conditions are presented and development requirements identified. Capital and operating costs are presented for the 50 MW demonstration plant and for commercial plants with capacities from 50 to 500 MW.

Work was also carried out to develop process design data and criteria. Experimental tests were carried out to identify candidate limestone sorbents (14 limestones were tested); to obtain data on sulfur removal, sorbent regeneration, and spent stone processing (five process concepts were tested); and to assess the environmental impact of the spent sorbent.

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[*] Esso Research Centre, Abingdon, England (Esso) refers to the process as the chemically active fluidized bed (CAFB) process.

[†] Watts are used to express electrical equivalent and Btu to express thermal equivalent throughout this report.

Conclusions and recommendations are based on the results of the work performed under this contract and on the projected costs and performance of alternative systems for sulfur oxide emission control which are available from outside sources. The systems considered to be alternatives to the CAFB process are stack-gas desulfurization via wet scrubbing and hydrodesulfurization (HDS) of residual fuel oil, including vacuum resid. Unfortunately, comprehensive cost and performance data on these systems are limited. The stack-gas desulfurization costs and performance are based on the recent TVA report by McGlamery and Torstrick.[3] The HDS costs and performance are based on the Cities Service/Hydrocarbon Research H-Oil process. The conclusions and recommendations for development of the CAFB process are dependent on the validity of cost and performance data available on these alternative processes. The Westinghouse assessment of the TVA cost estimates for stack scrubbers and the SWEC estimates for CAFB installations is that SWEC has incorporated a substantial element of conservatism in terms of overdesign (25 percent excess capacity is common in the SWEC design) and in estimating.(Piping, foundations, structural supports, etc. are estimated with very liberal cost factors for labor productivity, escalation,and contingency amounts.) This conservatism is not included in the TVA estimates. For the actual conditions to be found in real installations, the SWEC costs are believed to be more representative than the TVA costs by a factor of approximately 30 percent.

CONCLUSIONS

The primary conclusions are:

- There will be sufficient quantities of high-sulfur residual fuel oils,and there are existing power plants such that a market exists for retrofitting electrical utility power plants to utilize high-sulfur fuel oils.
- The process offers the potential to utilize low-grade petroleum or synthetic fuel fractions with high metals content which cannot be economically utilized in

alternative processes for producing clean power with conventional boilers.

- On the basis of the development work which has been carried out, a 50 MW plant can be built.
- The economics of the CAFB process are sufficiently attractive that the process merits development for utilizing low-grade, high metals content, residual fuel oils.
- A pressurized fluidized bed residual oil gasification process for firing a combined cycle power plant is lower in cost than the atmospheric-pressure retrofit system. The total power plant cost for the pressurized power plant system will be significantly lower than for a conventional power plant.

Conclusions from the assessments of the market, technology, process economics, and the demonstration plant design and cost estimate are given below.

Market

- Atmospheric-pressure operation of the CAFB process is most applicable to the electrical utility industry as a boiler retrofit for oil- or gas-fired boilers. The atmospheric-pressure process is not generally attractive for new boilers or retrofit of coal-fired boilers because of the current trend toward coal. The largest market is for boiler sizes ranging from 50 to 400 MW.
- Vacuum bottoms or other low-grade high metals content fuels are the most likely to be available for the CAFB process.
- Limestone availability for CAFB may be more restricted than for slurry scrubbing processes due to more stringent requirements on sorbent physical properties with CAFB.

- Potential markets for CAFB by-product/waste stone are uncertain.

Technology

- Environmentally, CAFB appears superior to lime and limestone slurry flue-gas scrubbing because of the reduced impact of nitrogen oxide, solid waste, and process water requirements.
- Higher fuel efficiencies are realized with CAFB than with HDS.
- The ability of CAFB to utilize low-grade petroleum or synthetic fuel fractions with high metals content provides the potential for a fuel source which may not be feasible with HDS or stack-gas cleaning processes.
- CAFB is in a much earlier state of development than is lime/limestone slurry scrubbing or HDS, although major development work is still required in slurry scrubbing processes to demonstrate reliability.
- CAFB has the potential to offer high reliability and lower space requirements than lime/limestone slurry scrubbing.
- It has not been established whether there is, in general, sufficient physical space at a majority of boiler plant sites for either CAFB or slurry scrubbing retrofit.
- The utilization of a clean fuel, such as from HDS, is the most convenient technological option available to the utility.

Economics

- The capital cost of CAFB is 25 to 50 percent greater than that of limestone scrubbing and is comparable to regenerative stack-gas cleaning costs.

- The fuel adder for CAFB is competitive with limestone scrubbing if low-grade fuels available to CAFB are 10 to 20¢/10^6 Btu cheaper than fuels suitable for firing in a plant with limestone scrubbing.
- HDS of vacuum bottoms with low metals content is not competitive, even though operating at a higher load factor, unless a unit larger than 25,000 bbl/d is built to supply more than three 200 MW boilers, or a unit larger than 35,000 bbl/d is built to supply two 500 MW boilers. Thus, HDS requires more immediate commitment of capital than would a single CAFB unit.

Demonstration Plant Design and Cost

- The design basis for the gasifier-regenerator system is based on the data obtained in the 750 kw pilot plant unit.
- Experimental tests were carried out to evaluate five spent stone processing options. These data indicate
  - Sulfur dioxide and spent limestone from the regenerator can be combined in a dry sulfation process to form calcium sulfate which is suitable for disposal. More than 90 percent sulfation was obtained in laboratory tests.
  - High temperature (1400°C) sintering of the spent sorbent from the regenerator is a feasible method for processing the spent limestone for disposal.
  - Slurry recarbonation of spent limestone flue gas will produce a material which is environmentally suitable for disposal and may be marketable.
  - Treatment of the spent limestone with dilute sulfuric acid will produce material containing

  primarily calcium sulfate (94% sulfation) which is environmentally suitable for disposal.
  - Oxidation of the calcium sulfide to calcium sulfate followed by high-temperature recarbonation of the calcium oxide is not practical for a regenerative process and will only be feasible in a once-through sorbent process if the residual calcium sulfide is not released.

- The concept chosen for the CAFB demonstration plant uses dry sulfation of the waste sorbent to eliminate the sulfur recovery step. This appears to be the most economical processing concept other than once-through operation with direct disposal of the utilized sorbent.
- Screening tests were carried out on 14 candidate limestones for the demonstration plant. Limestone 1359 and aragonite were selected as the candidate sulfur sorbents.
- High excess capacity factors have been applied to the demonstration plant design to assure flexibility of operation.
- The demonstration plant cost is about $7,400,000, including the fuel oil system. This cost does not include start-up cost, interest during construction, and engineering fee.
- The most important cost component in the plant is the hot fuel-gas piping system. Options to its present design should be considered. Other costly components are the gasifier cyclones and fines recycle system.
- Because of cost considerations the four fuel-gas burners should be placed in a single boiler face rather than tangentially in the four corners.

- A new induced draft (I.D.) fan should not be installed until demonstration plant operation indicates that it is required to meet full boiler capacity.

Development Requirements

A number of aspects of the demonstration plant require further development:

- The utilization of low-grade petroleum or synthetic fuel fractions with high metals content
- The pilot-scale demonstration of the dry sulfation spent stone processing system
- The assessment of the environmental aspects of stone disposal and utilization options
- The demonstration of the gasifier-to-regenerator solids transport system and sorbent pulverizers on commercial-scale equipment
- The review of safety requirements
- The development of alternative and advanced subsystems for the process—for example, utilization of impure limestone, alternative spent stone processing, once-through operation
- Plant layout and fuel-gas piping concepts.

RECOMMENDATIONS

On the basis of the market, design, and economic studies, and assessments carried out, Westinghouse recommends that:

- Detailed design and construction of the 50 MW demonstration plant utilizing regular-grade, high-sulfur residual oil not be carried out, because conventional technology exists which can utilize these fuels and which is cost competitive with CAFB on the basis of available costs.

- Development of the atmospheric-pressure fluidized bed residual oil gasification/desulfurization process be directed toward the utilization of low-grade, high metals content fuels, such as vacuum bottoms, which are not compatible with conventional boiler/stack-gas cleaning systems.
- Fuel availability for the process be investigated to assess market need and market potential for the process. Fuels to be investigated should include but not be limited to vacuum bottoms, Venezuelan bitumens, residual oil from shale, tar sands, and liquefaction of coal.
- Development work be directed toward the following critical areas:
  - Process demonstration with low-grade, high-metals content petroleum fuels in the Esso CAFB pilot plant
  - Demonstration of dry process(es) for disposal/utilization of sulfur dioxide and spent stone in pilot-scale tests
  - Demonstration of the pulsed flow solids transport system between the gasifier and regenerator in a commercial-scale facility
- Detailed design and construction of a 50 MW demonstration plant utilizing low-grade, high metals content fuels be carried out if completion of the critical experimental tests identified and review of fuel projections demonstrate process operability and fuel availability.
- A program be initiated to assess the application of a pressurized fluidized bed residual oil gasification/combined cycle power plant system to utilize regular high-sulfur residual oil and low-grade, high metals content fuels.

II. INTRODUCTION

Achievement of the national goal of fuel resource independence rests largely on the success of massive development programs in coal and nuclear fuel utilization. It is clear that petroleum derivatives will continue to supply a large portion of the United States' electric utility fuel demand during the years before national energy independence is realized. Two options presently exist for the environmentally sound utilization of atmospheric residual fuel oils for electric power generation: conversion to low-sulfur fuel oils (hydrodesulfurization) at the refinery or stack-gas cleaning (limestone slurry scrubbing, lime slurry scrubbing, and others) as a utility boiler retrofit or new boiler feature. In most cases the utilization of refinery vacuum bottoms as a fuel oil is not economically feasible with these two options because of their high metals content. The goal of the fluidized bed residual oil gasification process is to provide a third alternative which utilizes heavy fuel oils (atmospheric residual or vacuum bottoms) to generate clean power in terms of gaseous emissions and liquid and solid wastes.

The Environmental Protection Agency (EPA) through their Office of Research and Development (ORD) is funding the development of an atmospheric-pressure fluidized bed oil gasification process for firing in a conventional power plant. A simplified illustration of the process concept is shown in Figure 1. The Esso Petroleum Company, Ltd., Abingdon, England (Esso) invented the chemically active fluidized bed (CAFB) process and is carrying out pilot-plant studies. Residual fuel oil is added to a simple, open, fluidized bed of limestone with sufficient air—to 25 percent stoichiometric—to maintain the bed at about 870 to 970°C (1600 to 1778°F) and to react the oil to produce a hot fuel gas. A summary of the Esso development work is shown in Table 1. With EPA funding Esso has operated a 750 kw CAFB pilot plant more than 2600 hours and has provided design and operating information required for the development

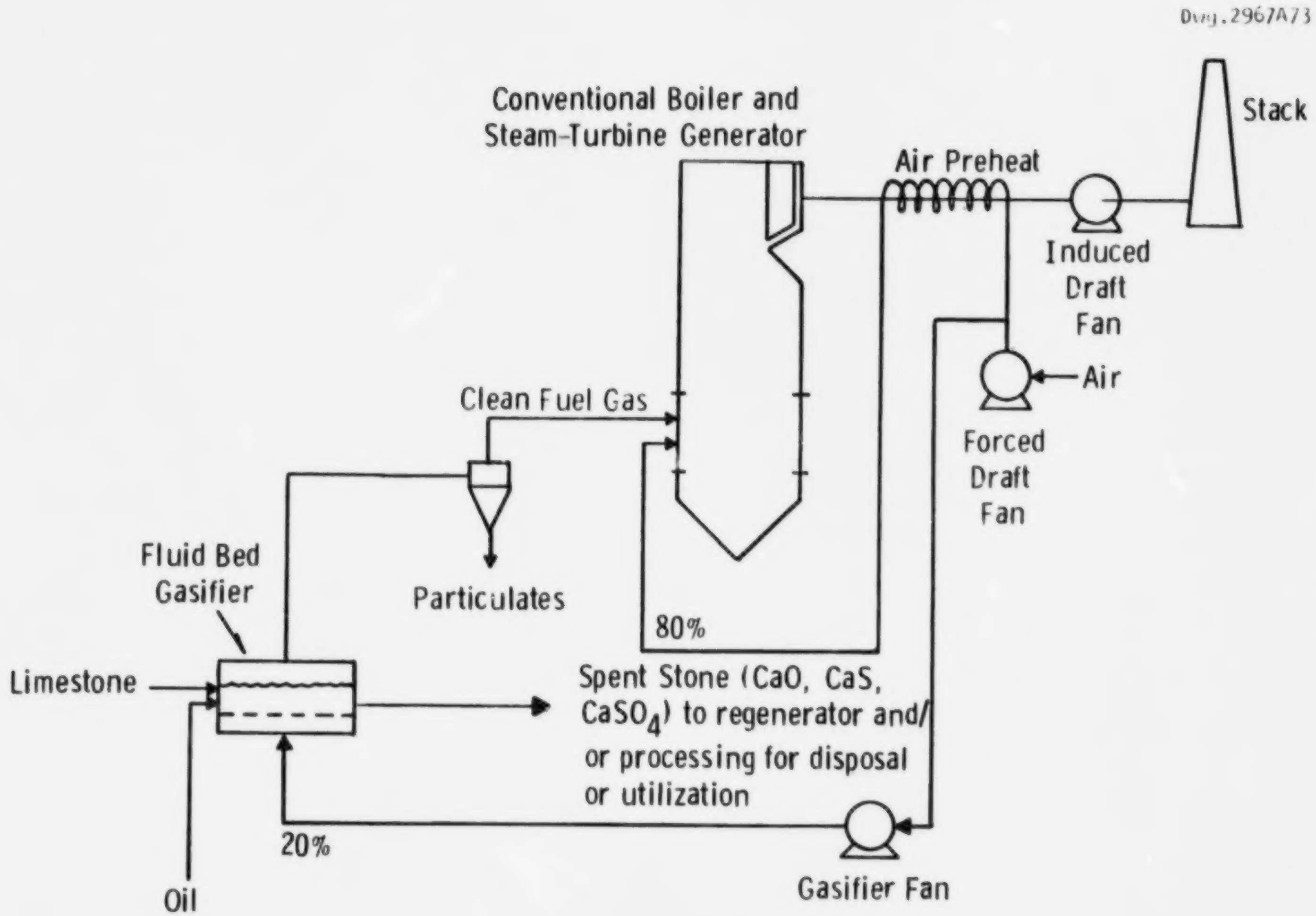

Dwg.2967A73
Conventional Boiler and Steam-Turbine Generator
Stack
Air Preheat
Induced Draft Fan
Air
Forced Draft Fan
Clean Fuel Gas
Fluid Bed Gasifier
Particulates
80%
Limestone
Spent Stone (CaO, CaS, CaSO4) to regenerator and/ or processing for disposal or utilization
20%
Oil
Gasifier Fan

Figure 1—Atmospheric fluidized bed oil gasification power plant

program. EPA contracted with Westinghouse in 1969 to carry out commercial plant conceptual designs and an evaluation of the CAFB process.

Table 1

PHASES OF CAFB EXPERIMENTAL DEVELOPMENT

Phase	Accomplishments	Time period
Pre-EPA	Feasibility in batch units	1966-1970
EPA Phase I	• Comparison of fuels and limestones in batch units • Intensive process variables study in batch units • Design, construction, and commissioning of continuous pilot plant	June 1970-March 1972
EPA Phase II	• Comparison of fuels and limestones in batch units • Study of process variables, design alternatives, and process problems in continuous pilot plant	July 1972-Dec. 1973
EPA Phase III	• Support for demonstration plant design • Investigation into trace element retention	July 1973-Jan. 1975

In November 1972, New England Power Service Company of New England Electric System (NEES) agreed to participate in the program, and boiler No. 12 at the Manchester Street Station, Narragansett Electric Company, Providence, R.I. was selected as the demonstration plant site.[2]

A three-phase demonstration plant program has been conceived:

- Phase I. Preliminary design and cost estimate of demonstration plant installed on an existing boiler, followed by an assessment by all parties and a decision to stop or proceed

- Phase II. Detailed design and construction of demonstration residual oil gasification process
- Phase III. Developmental operation of the gasification process and integrated power plant.

Westinghouse, as prime contractor to EPA, subcontracted with Stone & Webster Engineering Corporation (SWEC) for the engineering services for Phase I. Combustion Engineering, Inc. (CE), was engaged in a boiler retrofit and burner evaluation study.

Objectives for the work performed under this contract included

- Assessment of the market for the process
- Selection of complete, integrated fluidized bed oil gasification/desulfurization/power plant concept
- Preparation of a preliminary design and cost estimate for the 50 MW demonstration plant
- Preparation of a conceptual design and cost estimate for a 200 MW plant
- Specification of a work scope required to carry out the demonstration plant program
- Assessment of alternative technology in light of the prior tasks

The results of the work carried out to meet these objectives are reported in this two-volume report.

III. PROCESS FUNDAMENTALS

The basic fluidized bed residual oil gasification or CAFB process includes four processing steps:

- Oil gasification/desulfurization
- Lime regeneration
- Waste stone processing
- Sulfur recovery (optional).

A number of variations of these processing steps have been considered in process evaluations[1,2]—for example, a once-through limestone mode of operation where the sorbent is highly utilized in the gasifier up to 90 percent and the sorbent regeneration step is eliminated. These concepts are illustrated in Figure 2. The regenerative mode has been selected as the demonstration plant basis and for commercial assessment of the CAFB process because it has been successfully tested on a pilot-plant scale (∿ 1 MW), and it permits operating with the minumum quantity of sorbent. A number of variations involving waste stone processing and sulfur recovery are still under active investigation. A dry sulfation process which contacts the waste stone with the regenerator sulfur dioxide product-gas and eliminates the sulfur recovery step has been selected for the demonstration plant preliminary design and commercial assessments.

The gasification/desulfurization step has been previously discussed in some detail.[4,5] A summary of the gasifier conditions, reactions, and performance is shown in Table 2. Residual oil (atmospheric or vacuum bottoms) is injected into a bed of lime (870°C/1600°F) along with 1/2 to 1-1/2 times the stoichiometric amount of makeup limestone and about 17 to 25 percent of the air required for complete combustion of the oil. The oil cracks, carbon is deposited on the lime particles, the fresh limestone calcines, and hydrogen sulfide (and other sulfur

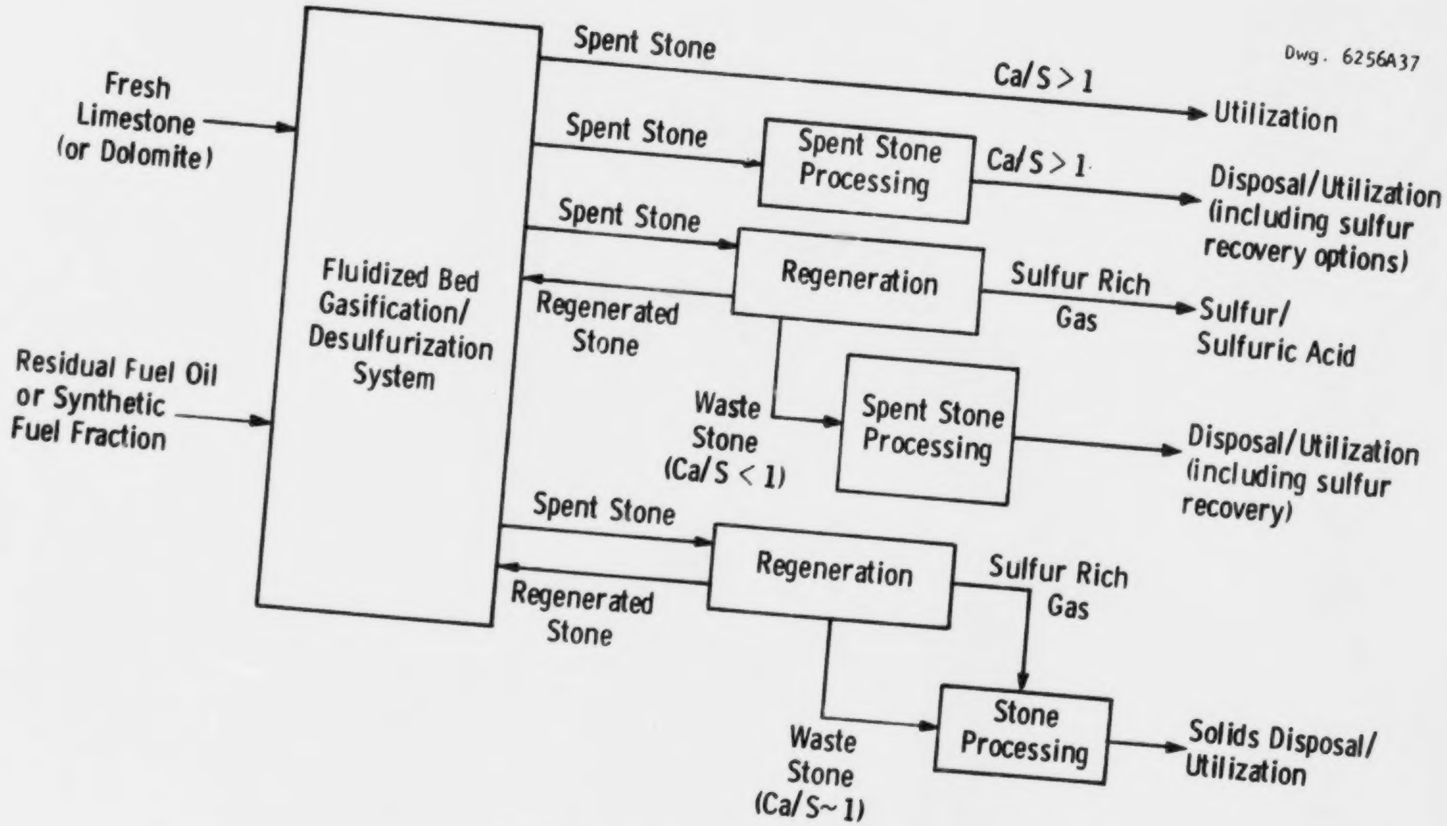

Dwg. 6256A37
Fresh Limestone (or Dolomite)
Residual Fuel Oil or Synthetic Fuel Fraction
Fluidized Bed Gasification/ Desulfurization System
Spent Stone
Ca/S > 1
Utilization
Spent Stone
Spent Stone Processing
Ca/S > 1
Disposal/Utilization (including sulfur recovery options)
Spent Stone
Regeneration
Regenerated Stone
Sulfur Rich Gas
Sulfur/ Sulfuric Acid
Waste Stone (Ca/S < 1)
Spent Stone Processing
Disposal/Utilization (including sulfur recovery)
Spent Stone
Regeneration
Regenerated Stone
Sulfur Rich Gas
Waste Stone (Ca/S~ 1)
Stone Processing
Solids Disposal/ Utilization

Figure 2— Sulfur removal system process concepts

Table 2

SUMMARY OF GASIFIER CONDITIONS AND PERFORMANCE

Conditions

- Temperature - 870 to 970°C (1600 to 1778°F)
- Pressure - atmospheric
- Air/fuel ratio - 17-25% stoichiometric
- Fresh limestone rate - 0.5-1.5 times stoichiometric
- Bed sulfur content - 4-6 wt % sulfur as CaS
- Bed carbon content - 0.1-0.3 wt % carbon
- Fluidization velocity - 1.2-1.8 m/sec (4-6 ft/sec)
- Bed depth - 0.6-1.2 m (2-4 ft)

Reactions (two-region fluidized bed)

Oil thermal cracking → Carbon + H_2 + hydrocarbons + H_2S, CS_2, COS

Oxidizing region

$C + \{O_2\} \rightarrow CO_2, CO$

$CaS + 3/2\ O_2 \rightarrow CaO + SO_2$

$CaS + 2O_2 \rightarrow CaSO_4$

$\mathrm{Oil} + O_2 \rightarrow CO_2, H_2O$

Reducing Region

$CaCO_3 \rightarrow CaO + CO_2$

$H_2S + CaO \rightarrow CaS + H_2$

$CaSO_4 + \mathrm{hydrocarbons} \rightarrow CaS$

Performance - demonstrated in batch tests and pilot-plant runs[4]

- 90% sulfur removal or greater
- Removes 100% of fuel vanadium, 75% nickel, 20% sodium
- Produces fuel gas having hot heating value 7.450 kJ/m^3 (200 Btu/scf or greater)
- Combustion of the fuel gas possible with low excess air: acts as two-stage combustor which yields reduction in nitrogen oxides (from 270 ppm to 155 ppm on pilot plant)
- High reliability demonstrated
- Limestone makeup rates down to 0.5 of stoichiometric are possible-nominally operates at stoichiometric makeup in the regenerative system.
- Carbon deposition (tars) and lime accumulation may occur in the fuel-gas lines and cyclones.

compounds) are released. The deposited carbon, and some fuel oil, is combusted while the hydrogen sulfide reacts with the lime to form calcium sulfide. A number of other side reactions also occur[4,5] and are discussed in Appendix D. The gasifier temperature is controlled by the air/fuel ratio and either stack-gas recycle, water, or steam injection. Control not only of sulfur dioxide but also of nitrogen oxides and fuel oil metals is realized.

The details of the regeneration chemistry have been previously described[4,5] and are summarized in Table 3 along with the regenerator performance. The regenerator chemistry is also discussed in Appendix D. The calcium sulfide produced in the gasifier is circulated to the regenerator and is contacted by air at 1050 to 1100°C (1922 to 2012°F) to generate calcium oxide, some calcium sulfate, and a sulfur dioxide gas of about 8- to 10 volume percent sulfur dioxide. Deposited carbon is also combusted. The regenerated sorbent is recycled to the gasifier. The rate at which the sorbent is circulated between the gasifier and regenerator controls the regenerator temperature. The sulfur dioxide is at a concentration suitable for sulfur recovery or sulfuric acid production.

A number of processes are being considered for converting the spent solids from the regenerator into an environmentally acceptable form for disposal or by-product utilization. The regenerator material is about 92 wt % calcium oxide, about 3 wt % calcium sulfide, about 3 wt % calcium sulfate, and the remainder inert compounds. The technical support work to develop these processes is contained in the appendices.

Examples of options which might be utilized, depending on market conditions and power plant location, are dry sulfation of the waste with the regenerator sulfur dioxide (Appendix L), dead-burning of the waste solids (Appendix O), dry oxidation of the calcium sulfide followed by recarbonation of the waste or silica sintering (Appendix G), all of which produce a dry solid product. The options of wet carbonation of the stone (Appendix M) or wet sulfation (Appendix N) have also been considered but are considered less attractive. The waste solids may also be valuable without further processing for their calcium oxide or their vanadium content, which may be about 1 wt % in some cases.

Table 3

SUMMARY OF REGENERATOR CONDITIONS AND PERFORMANCE

Conditions

- Temperature - 1050-1100°C (1922-2012°F)
- Pressure - atmospheric
- Fluidization velocity - 1.2-1.8 m/sec (4-6 ft/sec)
- Bed depth - 0.61-1.22 m (2-4 ft)
- Bed sulfur content - 0.5 - 1 wt % sulfur as CaS
- Bed carbon content - 0% carbon
- About 0.1 mole % oxygen in product gas

Reactions

$C + O_2 \rightarrow CO_2$

$CaS + 3/2\ O_2 \rightarrow CaO + SO_2$

$CaS + 2O_2 \rightarrow CaSO_4$

$1/4\ CaS + 3/4\ CaSO_4 \rightarrow CaO + SO_2$

Performance[4,5]

- SO_2 content of product gas - 8-10 mole percent
- $CaSO_4$ production in regenerator - amount produced depends on stone carbon content, operating temperature, excess oxygen level

IV. ASSESSMENT

MARKET

Perspective

A preliminary review of the total market for oil gasification plants indicates that the majority of plants range between 50 to 600 MW for both new and retrofit applications and can be used in either the utility or industrial markets. The advantages of economy of scale may become limited by the physical size of the units, thus requiring a modular design for the large-capacity plants. Current trends toward energy conservation have led to reduced income for utilities as well as delays in power generation expansion across the country. This consequence, the projected utilization of coal, the potential for large (800 to 1400 MW) nuclear-powered stations, and the uncertain petroleum availability for power generation could minimize the application of oil gasification for new installation. If current delays in power generation expansion continue, however, there could be a market for a pressurized fluidized bed residual oil gasification/combined cycle power plant concept.

Coal-fired units have been eliminated from this analysis. It is believed that the current shift to coal and resources conservation ethics will preclude any significant future conversion of coal units to oil. The retrofit market for existing oil- or gas-fired units is, therefore, considered to be primary. A summary of existing oil- or gas-fired boilers of commercial size is shown in Table 4. On the basis of this survey, it would appear that a 200 MW unit would be a good choice for modular design. The total potential application is approximately 200 in number. This represents a much smaller market segment when compared to the number of potential 0 to 100 MW applications. On the basis of these considerations, a 200 MW unit has been selected for commercial comparison.

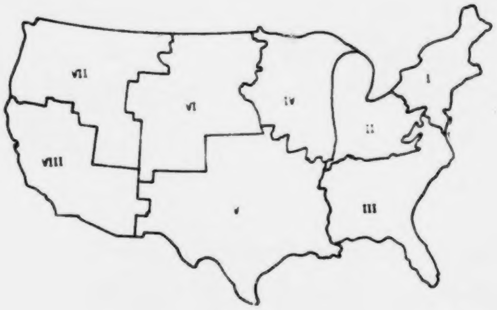
Table 4

SUMMARY OF EXISTING OIL- OR GAS-FIRED BOILERS[a]

FPC Region[b]	Unit capacity (MW)			
	0-100	101-200	201-400	400-1000
1	167	33	13	7
2	Mostly coal-fired			
3	54	17	13	5
4	Mostly coal-fired			
5	197	50	35	29
6	Mostly coal-fired			
7	Mostly coal and hydro			
8	53	18	8	6
Total[c]	471[d]	118	69	47[e]

[a] Considers boilers whose primary fuel is oil or gas regardless of the unit design fuel.

[b] The Federal Power Commission (FPC) regions are illustrated in the following outline map of the United States:



[c] If all of the boilers which could be fired with coal were eliminated from consideration in the tables, Region 1 would be effectively reduced to no boiler candidates, the number of candidates in Region 3 would be reduced by one-third, and Regions 5 and 8 would be essentially unaffected.

[d] Essentially all the boilers smaller than 50 MW are 20 to 40 years old.

[e] Of the units between 400 and 1000 MW, 11 are greater than 600 MW and 2 are greater than 800 MW.

The specific application of the oil gasification process will need to consider other factors in addition to those mentioned. The process economics are influenced by the power station's load factor, utility costs, space availability, proximity to refineries, and so on. The distribution of fuel is an important factor for small plants and for plants which would utilize low-grade fuels such as vacuum bottoms. The availability of raw materials may also influence the decision to utilize this process. The disposability of the by-product material may be either an asset or a liability, depending on the specific application. Thus, each potential application must be fully evaluated for its individual merits.

The major market areas are located on the East, West, and Gulf Coasts. The market is almost equally distributed in these areas.

Market Factors

Environmental Standards

Elimination or relaxation of current emission standards would permit the direct use of high-sulfur fuels and reduce the trend for increased use of low-sulfur fuels. Thus, the emphasis on developing various pollution abatement apparatus is also diminished. While a reduction in localized air pollution requirements may be considered for brief periods, long-range national programs are directed toward overall improvement of air quality standards.

Environmental Legislation

The continually increasing interest in utilizing existing coal resources may prompt legislation restricting the use of alternative fuels for specific applications. The value of crude oil feedstocks and their derivatives may result in legislation prohibiting their use for power generation. It is considered more likely, however, that oil feedstocks will be limited by supply. Thus, power utilities would be forced to use coal or refinery vacuum bottoms for fossil units. Vacuum bottoms, while oil derivatives, represent the end product of the fuel distillation process.

Feedstock Availability

The availability of high-sulfur fuel feedstocks is critical to the marketability of the residual oil gasification process. An assessment of various feedstocks (Table 5 and Appendix A) results in the conclusion that vacuum tower bottoms or other high metal content fuels are the most likely to be economically attractive for the CAFB process. The utilization of these low-grade petroleum residues is, unfortunately, unclear. Several processing alternatives have been proposed, including blending with low-sulfur and/or low-metals residual oils.

Commercial processes are being developed to desulfurize fuel oil from many crude oils. The cost of desulfurizing high-sulfur, high metal content residual oil fractions, however, is currently too high to permit the production of a fuel economical for power generation.

What quantity of the low-grade fuel would be available for utility use with the CAFB is not clear; however, quantities could be available to meet approximately 25 percent of the present oil-fired power generation capacity (Appendix A).

The ability of the CAFB process to handle very heavy high metal content resids is important. Large quantities of fuel liquids could be produced from such sources as the Alberta tar sands, the Orinoco tar belt, oil shale deposits, and coal deposits. The high boiling residues (+565°C/1050°F) of these "syncrudes" would be suitable for industrial and/or utility application by utilizing the CAFB process. These fuels could significantly increase the available feedstocks.

Feedstock Availability - Limestone

Limestone is generally available throughout the country.[6,7] While all kinds may be suitable to some extent, the process may require some modifications in order to handle varied particulate loading levels because of different limestone attrition rates. Thus, process economics are influenced by the rate of use of limestones based on their performance. Results of tests of various limestones from all sectors of the country

Table 5

FEEDSTOCK SELECTION

Feedstock	Comment
High-Sulfur Crude	Unlikely to be available for power generation use as part of a resources conservation program because of the valuable by-products which it contains.
High-Sulfur Residual	Foreign Currently 23% of our residual imports have high-sulfur content (Appendix A). While more widely available from European refineries because of their operational technique (up to 40% of every barrel of crude processed becomes resid), this report has conservatively precluded their use, assuming a limited import policy. Domestic While high availability is forecast, this report precludes their use, assuming restrictive legislation requiring them to be processed to vacuum bottoms, also part of the resources conservation program (Appendix A).
High-Sulfur Vacuum Bottoms	Foreign Unlikely because of current refinery operation mode and transportation economics; added high costs and equipment modifications would make its importation impractical. Domestic Most likely alternative based on availability, see Appendix A.

are provided in Appendix Q. In general, availability is no problem; limestones, however, must be selectively identified in each area for a specific application.

Competing Technology

Competing technology or the advent of advanced technological development resulting in a more practical and economical method of desulfurizing fuel would shift the problem of pollution control from the power utility to the fuel supplier. Some of these aspects are reviewed in the technology assessment below.

TECHNOLOGY

Technological aspects of the CAFB process are assessed and compared with two competing technologies for fuel utilization and pollution control -- lime/limestone slurry flue-gas scrubbing and hydrodesulfurization of residual fuel oils. The environmental impact, the present state of development, and the applicability of the technologies to the electrical utility industry are reviewed.

Environmental Impact

The environmental impact of the CAFB process and lime/limestone slurry flue-gas scrubbing are compared in Table 6. Details of CAFB environmental performance are presented in Appendix G. The basis of comparison is a 200 MW oil-fired power plant (retrofit). The CAFB process appears to be superior environmentally to limestone and lime slurry scrubbers. Sulfur removal capabilities for the processes are comparable, but the CAFB process provides a considerable reduction in nitrogen oxide emissions and consumes an order of magnitude less process water than do slurry scrubbing processes. Similar nitrogen oxide control could be achieved by the addition of special combustion equipment on the boiler. Auxiliary power requirements are comparable at three to four percent of plant capacity. The limestone usage is nearly identical for the processes -- about 1 mole of calcium per mole of sulfur removed from the fuel -- but the CAFB process

Table 6
ENVIRONMENTAL IMPACT COMPARISON

	200 MW conventional oil-fired power plant with limestone or lime slurry flue-gas scrubbing	200 MW conventional oil-fired power plant with CAFB oil gasification and dry sulfation of solid product
Plant Emissions		
SO_2 kg/GJ (lb/mm Btu)	0.15 (0.35)	0.15 (0.35)
NO_x kg/GJ (lb/mm Btu)[a]	0.34 (0.80)	0.17 (0.40)
Total Particulates kg/GJ (lb/mm Btu)	0.043 (0.10)	0.043 (0.10)
CaO	None	0.026 (0.06)
$CaSO_4$	None	0.0043 (0.01)
Solids Mg/day (tons/day)	145 (160)	145 (160)
chemical form[b]	$CaSO_3$, $CaSO_4$, $CaCO_3$, inerts (sludge)	> 95% $CaSO_4$, CaO, inerts
Liquids Mg (tons/day)[b]	145 (160)	None
chemical form	Water (sludge)	None
Resource Usage		
Limestone Mg (tons/day)	118 (130)	118 (130)
Process water, liters/day (gal/day)[c]	1.1×10^6 (3.0×10^5)	1.1×10^5 (3.0×10^4)
Auxiliary power (kw)[c]	$6 \times 10^3 - 8 \times 10^3$	$6 \times 10^3 - 8 \times 10^3$
Steam, kg/day (lb/day)[d]	3.2×10^5 (7×10^5)	None
Land, m^2 (acres)[b]	2×10^4-4×10^4 (5-10) per year per 3 m (10 ft) depth of pond	None
Ability to utilize atmospheric residual oils (#6 oil)	Yes	Yes
Ability to utilize vacuum bottoms with high metals content.	No	Yes

[a] NO_x emission for gasifier based on pilot-plant measurements and experience with fluidized bed combustion.

[b] Waste water based on 50% water sludge product. Typical sludge properties and land requirements are reported in J.W. Jones, "Environmentally Acceptable Disposal of Flue Gas Desulfurization Sludges: The EPA Research and Development Program" and C.N. Ifeadi, H.S. Rosenberg, "Lime/Limestone Sludge Disposal - Trends in the Utility Industry"; both presented at the EPA Flue Gas Desulfurization Symposium, Atlanta, Georgia, Nov. 1974.

[c] Process water and auxiliary power for limestone and lime slurry scrubbing based on G.G. McGlamery and R.L. Torstrick, "Cost Comparisons of Flue Gas Desulfurization Systems," presented at the EPA Flue Gas Desulfurization Symposium, Atlanta, Georgia, Nov. 1974.

[d] Based on steam for stack-gas reheat.

could potentially reduce this consumption to half that level by utilizing a regenerative operation with sulfur recovery. The CAFB process also produces a dry product with potential market value rather than a sludge which is difficult to handle and requires large land areas for disposal ponds. CAFB also permits the utilization of high-metals vacuum bottoms as a fuel. Additives exist which may permit a boiler to utilize these high metal fuels directly, with reduced boiler deterioration.

Hydrodesulfurization (HDS) requires the production of hydrogen and oxygen, which contributes to a substantial energy inefficiency. Substantial amounts of process water and cooling water are also used. HDS produces elemental sulfur as a by-product, but its market value is uncertain. Nitrogen oxide emissions would be reduced to some extent in the power plant using a hydrodesulfurized fuel oil without the addition of nitrogen oxide control equipment to the boiler burners. The ability of HDS to utilize vacuum bottoms economically is currently in doubt because of the effect of high metal content of the oil on the catalyst performance.[8]

Present State of Development

The status of the development of the three alternative technologies is considered in Table 7. CAFB is in the pilot-plant stage with extensive testing with atmospheric residual oil already completed. Vacuum bottoms utilization has been investigated on batch-scale equipment. CAFB stone processing techniques have been investigated only on laboratory-scale apparatus. Lime and limestone slurry scrubbing represents the most advanced of the flue-gas scrubbing technologies.[9] Regenerative flue-gas cleaning techniques are in stages of development similar to CAFB. Although lime/limestone scrubbing is near commercial status, it has received only limited utility acceptance. HDS represents commercial technology, although the utilization of vacuum bottoms is economically beyond the current limits of the technology.

Development areas exist for all of the competing technologies, but CAFB, by the nature of its development, has the widest range. None of the CAFB development areas are believed to be critical to technological

Table 7

PRESENT STATE OF DEVELOPMENT

	CAFB	Lime/limestone scrubbing	HDS
Status	• Pilot plant (atm. residual oil utilization) • Batch units (vacuum bottoms utilization) • Laboratory units (stone processing)	• Near commercialization • Pilot plant (sludge processing)	• Commercial
Major Development Areas	• Stone processing/ sulfur recovery • Vacuum bottoms demonstration • Reduced limestone requirements • Control of coke deposition • Stone circulation • Detailed design and operation of demonstration plant	• Control of erosion, corrosion, plugging, scaling • Sludge disposal • Retrofit structural problems • Other mechanical problems such as bypass dampers	• Improved operation with vacuum bottoms

success. Lime/limestone slurry scrubbing processes have been making progress in reliability, but further development and demonstration is required.[10, 11] At the present time, the ability of HDS to utilize vacuum bottoms appears uneconomical, although processing alternatives do exist.[8,12]

Utility Applicability

The degree to which any of the three alternative technologies meet the requirements of the utility industry depends on considerations such as:

- Process reliability and maintenance requirements
- Process control ability
- Process complexity
- Retrofit boiler performance and life
- Process safety
- Process space requirements
- Raw materials availability.

It is likely that the utility industry would favor the firing of a clean fuel (such as that from a HDS process) over retrofitting either CAFB or lime/limestone slurry scrubbing processes based on these considerations. HDS does not require the utility to adapt to new technology. CAFB and lime or limestone slurry flue-gas scrubbing appear comparable in these aspects, with CAFB possibly providing greater reliability, lower space requirements, and improved boiler life.

High reliability has been demonstrated by CAFB in the pilot plant, except for the complications associated with coke-lime deposits in the cyclone inlets. This appears to be a minor problem in the light of improved operating methods to reduce the coke deposits now being investigated in the pilot unit and the effect of larger scale equipment on the time between shutdowns. Large land usage is required for disposal ponds with lime or limestone slurry flue-gas scrubbing, but CAFB requires only limited storage of dry waste/by-product material. Some plants may not provide the physical space required for a retrofit by either process. CAFB removes much of the fuel oil metals (vanadium, nickel, and sodium) associated with boiler tube corrosion. The availability of suitable limestones may restrict the application of both CAFB and wet scrubbing processes (Appendix Q).

The ultimate choice of technologies must be made on the basis of these technological issues combined with complex and unstable non-technical issues such as:

- Process cost
- Fuel cost, availability, reliability of supply
- National policy and pollution standards.

Geographical location will also play a role in the relative merits of these technologies. If a reliable and cheap source of vacuum bottoms is available for CAFB, the utility may be placed in a more attractive market situation.

The utilization of a pressurized fluidized bed residual oil gasification process with current combined cycle power generation technology was previously evaluated.[2] Application of CAFB at higher pressures offers the potential for higher plant efficiencies (up to 50 percent power plant efficiency) than can be obtained with conventional power plants; the ability to utilize more economically low-grade, high-metals content residual fuel oils in new power plants, and the ability to reduce construction time over conventional plants for new installations. These factors are important enough in the overall energy and environmental effort to warrant further consideration of the pressurized concept.

Summary

Environmentally, CAFB appears superior to lime and limestone slurry flue-gas scrubbing because of the reduced impact of nitrogen oxide, solid waste, and process water requirements. Higher fuel efficiencies are realized with CAFB than with HDS. The ability of CAFB to utilize vacuum bottoms provides the potential for a fuel source which may not be feasible with HDS or stack-gas cleaning processes.

CAFB is in a pilot-plant stage of development while lime/limestone scrubbing is near commercial status and HDS represents commercial technology.

The applicability of the three technologies to the electrical utility industry is a complex consideration. Technologically, HDS provides many advantages to the utility operator. CAFB provides a number of potential advantages over lime/limestone scrubbing -- possibly improved reliability, lower space requirements, and improved boiler life. When unstable factors such as process cost, fuel cost, fuel availability, and national policy are considered, many of the technological issues may be relatively unimportant. The greatest potential advantage of CAFB operating at atmospheric pressure or pressurized for combined cycle power appears to be its ability to utilize vacuum bottoms and, thus, to provide a fuel source to the utility which might be otherwise unavailable.

ECONOMICS

An assessment of costs for several alternative means of sulfur oxide pollution control has been prepared. The most likely alternatives to fuel desulfurization via CAFB processing, such as vacuum resid HDS and stack-gas desulfurization via wet scrubbing have been reviewed. Capital and operating costs have been assembled for the Cities Service/Hydrocarbon Research H-Oil process, a limestone slurry process for stack-gas purification under development by Bechtel and others, and a catalytic oxidation (Cat-Ox) process for stack-gas purification under development by Monsanto.

H-Oil was chosen to represent oil HDS for two reasons:

- The fast percolating bed of catalyst used in commercial applications of the H-Oil process provides the possibility for continuous catalyst replacement in the desulfurization of high metals content vacuum resid.
- Recent data are available from a reliable source (Hydrocarbon Processing Magazine's 1974 Refining Handbook issue[13]) on costs for a commercial-scale vacuum resid processing installation.

The limestone slurry process was chosen to represent the lowest cost system for sulfur dioxide removal from stack gas, and the Cat-Ox process by Monsanto was chosen to represent a high-cost system for sulfur dioxide removal from stack gas. It should be noted that the high cost of the Cat-Ox process is partly due to the avoidance of sludge or waste stone disposal problems. Both limestone slurry and Cat-Ox process costs were obtained from a recent TVA/EPA paper, "Cost Comparisons of Flue Gas Desulfurization Systems" by McGamery and Torstrick.[3] A substantial effort was made to place all published costs on comparable bases, as will be discussed in some detail below.

Capital Costs

SWEC prepared capital cost estimates for two CAFB plants, a 50 MW demonstration plan (see Appendix F) and a 200 MW demonstration plant (see Appendix H). The tabulation in Table 8 presents these two estimates by SWEC as costs I (50 MW) and II (200 MW). The SWEC estimates have been modified to a slightly different form in Table 9, by showing the new burner costs broken out from the process equipment costs and identified as allowances. Other amounts have been added to the SWEC estimate for contingency, fee, start-up costs, and interest during construction. The percentages used for these added amounts were derived from the McGlamery and Torstrick paper mentioned above. Also allowances for new induced draft (I.D.) fan costs were added to the SWEC estimates. Thus, bare cost estimates by SWEC of $7,360,100 and $21,583,500 for single-train demonstration plants of 50 and 200 MW, respectively, have become total capital costs of $9,497,000 and $28,155,000 in Table 8.

A careful review of the SWEC cost estimates discussed above permits the following conclusions to be reached:

- The 200 MW estimate, as well as the 50 MW estimates in Table 8 by SWEC, are both based on a 20 percent air/fuel (A/F) ratio. The 200 MW 20 percent air/fuel ratio case (Appendix H) prepared by SWEC assumes 16 gas burners.
- The 200 MW estimate includes new limestone storage and ash storage silos. Since the 50 MW design used existing coal

and coal-ash storage facilities for fresh and spent stone, no costs for storage silos were included in that estimate. The 200 MW design includes costs for nearly three weeks' storage of fresh and spent stone.

- The 200 MW estimate includes compressor costs that were derived from the 50 MW estimate by direct multiplication by four.
- The 200 MW instrumentation materials cost is 2.7 times the cost of instrumentation materials in the 50 MW estimate.

New estimates using the low air/fuel ratio (17 percent) data presented in SWEC's flow diagram and material balance, Figure 3, were made in-house by Westinghouse on the basis of the facts presented above and on the SWEC cost detail summary. The new estimates proportioned costs for the low air/fuel case, 50 MW unit equipment, from the 20 percent air/fuel case, 50 MW equipment, by taking the ratio or equipment capacity to the 0.7 power, and multiplying costs by the resultant number. For example, the gasifier air and recycle gas (stream 2) in the low air/fuel case is 80 percent of the same stream in the base case. That capacity ratio, 0.8, to the 0.7 power is 0.855. The gasifier blower cost for the base case is thus multiplied by 0.855 to arrive at a cost for the low air/fuel case. New costs for the 50 MW and 200 MW units were derived in this way. At the same time, the 200 MW costs were also adjusted to reflect smaller stone storage silos (one week instead of three weeks' storage), to factor compressor costs from 50 MW to 200 MW using the 0.7 power instead of the direct multiplier SWEC used, and to factor instrumentation materials from a 50 MW to a 200 MW single-train plant using a 1.7 multiplier rather than the 2.7 multiplier SWEC used. The smaller multiplier is more appropriate for reflecting the same number and complexity of instruments in the large and small plants alike, but longer tubing and wire runs in the larger plant because of the physical size of the increased system. Thus, bare cost estimates of $6,648,000 and $17,252,000 for single-train, low air/fuel ratio plants of 50 and 200 MW, respectively, were obtained by Westinghouse estimates, and these become $8,528,000 and $22,305,000 total capital cost, as summarized in Table 8.

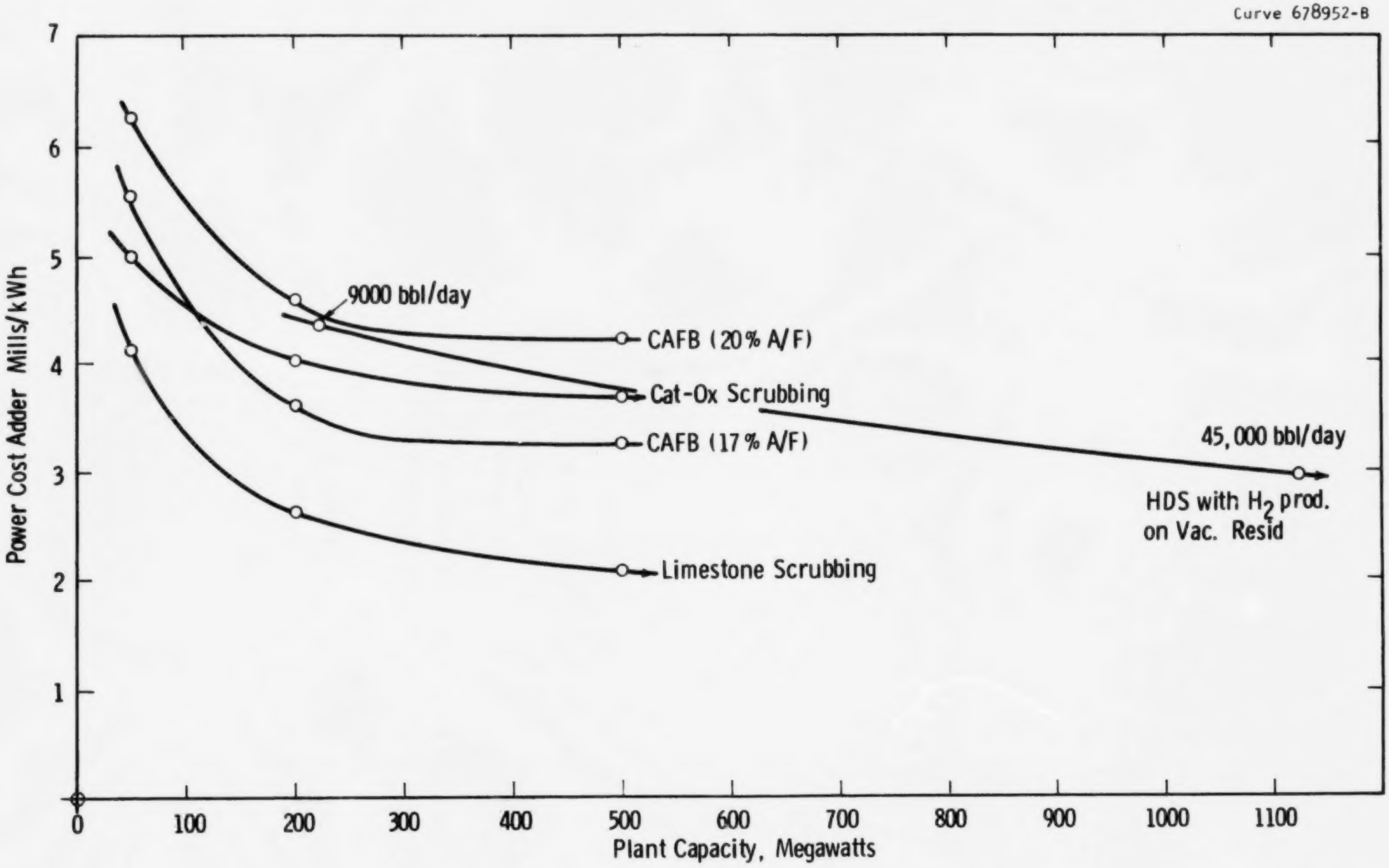


Figure 3 – Power cost adder for sulfur removal systems

To arrive at costs for CAFB units retrofitted to a 500 MW two-boiler, oil-burning installation, each 200 MW unit cost (one 20 percent air/fuel ratio, one 17 percent air/fuel ratio) was increased to 250 MW capacity using the 0.7 power, and then the cost was doubled to represent two complete trains of 250 MW each. The 250 MW CAFB gasifier would be approximately 13.7 m (45 ft) in diameter, approaching the size of the largest size cat cracker. The McGlamery and Torstrick paper also uses two 250 MW scrubbing systems for their 500 MW cost studies. It can be concluded that, both from a size standpoint and an equipment reliability standpoint, it is advisable to assume that a 500 MW capacity system would be retrofitted with two CAFB units as has been assumed in this economic assessment. The 500 MW CAFB total capital costs, as summarized in columns V and VI of Table 8, are $65,253,000 and $51,085,000 for the 20 percent air/fuel and 17 percent air/fuel cases, respectively.

The first process cost to compare with CAFB costs was the vacuum resid HDS H-Oil process. Cost and utility data available for the H-Oil system at 17,000 bbl/d permitted the evaluation of this process at two capacities bracketing the base point capacity (9000 bbl/d and 45,000 bbl/d). The 9000 bbl/d capacity of the small HDS unit approximates that required to fuel a 225 MW power plant, and the 45,000 bbl/d capacity of the large HDS unit is as large an installation as is practical to handle the resid from two or more two-stage crude oil stills of large capacity (160,000 bbl/d each) at a single site.

Based on H-Oil's $710 per daily barrel direct costs plus the hydrogen production system costs from a private communication with Stanford Research Institute, 1972,[14] and the use of a cost exponent of 0.65 derived for HDS systems by Guthrie of Fluor Coporation in a 1970 article in Chemical Engineering,[15] the total direct costs of the grass roots HDS units were $18,200,000 for the 9000 bbl/d installation and $50,400,000 for the 45,000 bbl/d installation ($2020/daily bbl and $1120/daily bbl, respectively). The total capital costs for those two units, as summarized in columns VII and VIII of Table 8, are $29,523,000 and $83,358,000. Due to the relatively established nature of the H-Oil process, half of the start-up

Table II
COMPARATIVE CAPITAL COSTS FOR SULFUR REMOVAL SYSTEMS

	I CAFB 50 MW unit (20% A/F est.)	II CAFB 200 MW unit (20% A/F, 1 train est.)	III CAFB 50 MW unit (17% A/F est.)	IV CAFB 200 MW unit (17% A/F ratio, 1 train low storage, etc.)	V CAFB 500 MW unit (20% A/F est.)	VI CAFB 500 MW unit (17% A/F est.)	VII Grass roots HDS unit with H_2 production (9000 bbl/d unit) (supply for 225 MW)	VIII Grass roots HDS unit with H_2 production (45,000 bbl/d unit) (supply for 1125 MW)	IX Limestone scrubbing 50 MW unit	X Limestone scrubbing 200 MW unit	XI Limestone scrubbing 500 MW unit	XII Cat-Ox process 50 MW unit	XIII Cat-Ox process 200 MW unit	XIV Cat-Ox process 500 MW unit
Process Equip. in Place	$ 2,378,800	$ 7,355,000	$ 2,146,000	$ 5,015,000	$ 16,496,000	$ 12,625,000	$ 7,761,000 a	$ 71,492,000 a	$ 1,087,000a	$ 3,254,000 a	$ 6,350,000 a	$ 1,392,000 a	$ 4,696,000 a	$ 10,386,000 a
Process Mat'l's & Labor	2,803,200	9,240,000	2,529,000	8,162,000	22,189,000	17,562,000	10,439,000 a	28,908,000 a	1,461,000a	4,377,000 a	8,542,000 a	1,873,000 a	6,315,000 a	13,969,000 a
Total Directs	$ 5,182,000	$16,595,000	$ 4,675,000	$13,177,000	$ 38,685,000	$ 30,187,000	$ 18,200,000	$ 50,400,000	$ 2,548,000	$ 7,631,000	$14,892,000	$ 3,265,000	$ 11,011,000	$ 24,355,000
Distributables	850,000	580,000	767,000	1,991,000	5,845,000	4,561,000	2,750,000	7,600,000	418,000	1,153,000 b	2,250,000	536,000	1,664,000 b	3,680,000
Subtotal	$ 6,032,000	$19,175,000	$ 5,442,000	$15,168,000	$ 44,530,000	$ 34,748,000	$ 20,950,000	$ 58,000,000	$ 2,966,000	$ 8,784,000	$17,142,000	$ 3,801,000	$ 12,675,000	$ 28,035,000
Indirect Costs	1,250,000	2,275,000	1,128,000	1,796,000	5,271,000	4,113,000	2,480,000	6,867,000	614,000	1,040,000 b	2,029,000	788,000	1,500,000 b	3,319,000
Total Bare Cost	$ 7,282,000	$21,450,000	$ 6,570,000	$16,964,000	$ 49,801,000	$ 38,861,000	$ 23,430,000	$ 64,867,000	$ 3,580,000	$ 9,824,000	$19,171,000	$ 4,589,000	$ 14,175,000	$ 31,354,000
Contingency	518,000	1,655,000	467,000	1,318,000	3,868,000	3,019,000	1,820,000	5,040,000	255,000	763,000	1,489,000	329,000	1,101,000	2,236,000
Fee	259,000	828,000	237,000	659,000	1,934,000	1,509,000	910,000	2,520,000	128,000	382,000	745,000	165,000	550,000	1,118,000
Total Process Invest't	$ 8,059,000	$23,933,000	$ 7,274,000	$18,941,000	$ 55,603,000	$ 43,389,000	$ 26,160,000	$ 72,427,000	$ 3,963,000	$10,969,000	$ 21,405,000	$ 5,078,000	$ 15,826,000	$ 34,708,000
New I.D. Fan c	50,000	200,000												
Burner Cost d	78,000	138,000	78,000	288,000	650,000	650,000	200,000 (cat.)	2,000,000 (cat.)				200,000 (cat.)	700,000 (cat.)	2,228,000 (cat.)
Total Invest	$ 8,187,000	$24,271,000	$ 7,352,000	$19,229,000	$ 56,253,000	$ 44,039,000	$ 26,360,000	$ 74,427,000	$ 3,963,000	$10,969,000	$21,405,000	$ 5,278,000	$ 16,526,000	$ 36,936,000
Start up Costs	655,000	1,942,000	588,000	1,538,000	4,500,000	3,523,000	1,054,000e	2,977,000e	317,000	878,000	1,712,000	528,000	1,653,000	3,674,000
Interest during Const.	655,000	1,942,000	588,000	1,538,000	4,500,000	3,523,000	2,109,000	5,954,000	317,000	878,000	1,712,000	[illegible]	1,322,000	2,939,000
Total Capital Costs	$ 9,497,000	$28,155,000	$ 8,528,000	$22,305,000	$ 65,253,000	$ 51,085,000	$ 29,523,000	$ 83,358,000	$ 4,597,000	$12,725,000	$24,829,000	$ 6,2[illegible]	$19,501,000	$43,549,000
$/kw	190	141	171	112	131	102	131	74	92	64	50	125	98	87

a Proportioned from total directs for CAFB 20% A/F 200 MW unit

b Adj. to CAFB/SWEC %

c Allowance

d SWEC allowance

e Half normal charge due to advanced stage of HDS development

Table 9
COMPARATIVE OPERATING COSTS FOR SULFUR REMOVAL SYSTEMS, $/yr^a

	I CAFB 50 MW unit (20% A/F est.)	II CAFB 200 MW unit (20% A/F, 1 train est.)	III CAFB 50 MW unit (17% A/F est.)	IV CAFB 200 MW unit (17% A/F, 1 train) low storage, etc.	V CAFB 500 MW unit (20% A/F est.)	VI CAFB 500 MW unit (17% A/F est.)	VII Grass roots HDS unit with H_2 production (9000 bbl/d unit) (supply for 225 MW)	VIII Grass roots HDS unit with H_2 production (45,000 bbl/d unit) (supply for 1125 MW)	IX Limestone scrubbing 50 MW unit	X Limestone scrubbing 200 MW unit	XI Limestone scrubbing 500 MW unit	XII Cat-Ox process 50 MW unit	XIII Cat-Ox process 200 MW unit	XIV Cat-Ox process 500 MW unit
Limestone or Catalyst	$ 50,000	$ 200,000	$ 50,000	$ 200,000	$ 500,000	$ 500,000	$ 356,400	$ 1,782,000	$ 50,000	$ 200,000	500,000	17,500	69,100	172,800
Labor & Superv'n to Operate	149,800	149,800	149,800	149,800	149,800	149,800	149,800	149,800	210,200	210,200	210,200	63,100	63,100	63,100
Steam	Neg.	Neg.	Neg.	Neg.	Neg.	Neg.	58,500	192,300	34,900	138,000	345,000	12,500	50,100	125,300
Water	Neg.	Neg.	Neg.	Neg.	Neg.	Neg.	12,800	64,200	2,000	8,900	20,000	2,500	10,000	25,000
Power	206,500	826,100	140,000	540,000	2,065,000	1,400,000	311,800	1,559,200	78,700	315,000	787,400	90,400	361,800	954,400
Maintenance	207,900	661,800	187,000	527,100	1,547,400	1,207,500	728,000	2,016,000	203,800	630,500	1,191,400	130,600	440,400	974,700
Labor Costs	27,000	27,000	27,000	27,000	27,000	27,000	45,600	45,600	45,600	45,600	45,600	48,000	48,000	48,000
Capital Charges	1,415,100	4,195,100	1,270,700	3,323,400	9,722,700	7,611,700	4,998,900	12,420,300	684,900	1,876,000	3,699,500	978,000	2,905,600	6,454,700
Plant Ovhd	118,100	332,900	100,800	252,700	757,800	554,900	257,300	809,400	134,900	264,500	519,900	69,400	194,700	478,000
Labor Ovhd	15,000	15,000	15,000	15,000	15,000	15,000	15,000	15,000	21,000	21,000	21,000	6,100	6,100	6,100
TOTAL	$2,188,900	$6,407,700	$1,940,300	$5,035,000	$14,784,700	$11,465,900	$6,934,100	$19,049,800	$1,465,600	$3,709,800	$7,340,000	$1,408,100	$4,149,100	$9,206,100
H_2SO_4 or Sulfur Credit							(939,700)	(1,403,300)				(47,100)	(188,400)	(471,000)
Fuel for Process Heat							1,778,600	8,698,200				417,800	1,671,100	4,177,900
Net							$7,773,000	$26,344,700				$1,778,800	$5,631,800	$12,913,000
Mills/kWh	6.25	4.58	5.54	3.61	4.22	3.28	4.36	2.96	4.13	2.65	2.10	4.97	4.02	5.60
¢ / 10^6 Btu	62.5	45.8	55.4	36.1	42.2	32.8	43.6	29.6	41.3	26.5	21.0	49.7	40.2	56.0

^aBasis: 7000 hr/yr, 2.5% S Oil, $4/ton Stone, 70¢/1000 lb STM, 1¢/kWh, 8¢/1000 Gal H_2O, 16.9%/yr Cap. Charges, $1.65/liter for Catalyst, $4/Ton of H_2SO_4, Oil @ $1.50/MM Btu for Reheat, Maint. @ 8%/yr Limes., 4%/yr CAFB, CAT-OX & HDS as % of Total Direct Investment, Labor @ $8/Manhour, Plant Ovhd. at 20% of O & M Costs, Labor Ovhd. at 10% of Direct Labor Costs expense percentage used for CAFB was used to arrive at start-up costs for the HDS units. Contingency, fee, and interest during construction percentages remained the same for HDS and CAFB estimates.

Both limestone and Cat-Ox processes for stack-gas cleanup were assessed using the previously mentioned work of McGlamery and Torstrick for TVA/EPA. The costs given in their paper for scrubbers applied to 200 MW and 500 MW oil-fired units burning 2.5 percent sulfur oil were used as a basis for comparison with CAFB. The mid-'72 start-of-engineering cost basis was escalated to early-'75 start-of-engineering by using a 1.25 cost multiplier derived from Chemical Engineering's cost index. Also, the new unit 200 MW costs given in the TVA study were factored to retrofit cost bases by applying cost adders to the new unit costs derived from 500 MW studies of new and retrofit cost differences for oil-fired systems. These cost adders take into account that a new unit is designed for a 30-year life versus a 25-year life for a retrofit design and, more significantly, that a new unit does not have the costs associated with installing scrubbers in an existing plant with limited space and many existing interferences that must be moved to accommodate the retrofit scrubbers.

In order to compare capital cost estimates fairly to CAFB and HDS costs, the scrubbing process costs used were the total direct costs, and the same percentages were applied to those total directs as applied for the CAFB and HDS units. Thus, distributables (in other words, construction management and overhead costs) were applied at about 15 percent of total directs, as SWEC had done, instead of the 11 percent to 13 percent used in the TVA study. Also, indirect costs (engineering, procurement services, and so on) were applied at about 13.5 percent of total directs as SWEC had done instead of the 9 to 11 percent used in the TVA study.
All other costs, such as contingency, fee, and so on, were also uniformly applied on the same basis as in the TVA study. The 200 MW and 500 MW costs in $/kw were plotted in semilog paper as a function of capacity in megawatts and were extrapolated to give $/kw for 50 MW units. These extrapolated costs were then summarized in Table 8, along with costs derived directly from the TVA study, described above. Limestone scrubbing total

capital costs for 50 MW, 200 MW, and 500 MW installations, as summarized in columes IX, X, and XI of Table 8, are $4,597,000, $12,725,000, and $24,829,000, respectively. Cat-Ox scrubbing total capital costs for 50 MW, 200 MW, and 500 MW installations, as summarized in Columes XII, XIII, and XIV of Table 8, are $6,228,000, $19,501,000, and $43,349,000, respectively. All of the costs discussed above are subject to additional factors that could, during the construction and start-up of these facilities, be responsible for as much as doubling the total capital charges summarized above. As pointed out in the TVA study, premium labor charges to cut the construction schedule or to obtain adequate manning, plus special R&D costs for first of a kind units, plus cost escalations at higher rates than those discussed above, unusual equipment delivery delays, any provisions for spare equipment to increase reliability, and so on could -- and usually do -- make actual installations much more expensive than those in Table 8. Those added costs can be applied uniformly to all costs tabulated, however, and should not affect the comparative economic process assessment.

Pressurized oil gasification for firing in a combined cycle power plant is estimated to be lower in cost than the atmospheric-pressure retrofit system cost because of:

- More compact vessels and fuel-gas piping at 1520 bars (15 atm) pressure
- Shorter, less complex fuel-gas piping system
- Higher plant efficiency, which permits lower limestone and oil rate per MW of plant capacity
- The regeneration process which generates hydrogen sulfide, permitting elemental sulfur recovery
- The plant layout which is designed for low investment and maintenance.

The gasification process (including oil feeding, limestone handling, regeneration system, waste stone processing system, waste stone handling system, sulfur recovery process, and gasification system) for a 260 MW combined cycle plant is estimated to cost about $60 kw (direct cost) based on end-of-1974 costs. A low-cost combined cycle plant (consisting of two Westinghouse 501B gas turbines and a single 0.7 m (28.5 in) steam turbine provides a total oil gasification combined-cycle power plant capital cost of about $250/kw compared to about $400/kw for a conventional oil-fired power plant with stack-gas cleaning. This large cost difference is credited to the low cost of the combined cycle plant -- maximum shop fabrication and short construction time.

Operating Costs

The TVA/EPA study was used as the basis for operating cost comparisons. It was felt that this study, using power cost at 10 mills/kWh steam at 70c/454 kg (1000 lb), potable water at 8c/3785 liters (1000 gal), 14.9 percent per year for capital based charges, and so on was a better basis for comparison of plants in the 200-to-500 kw range than the operating costs suggested by NEES for the Manchester Street Station and used by SWEC (Appendices E and H). Table 9 summarizes the annual cost so calculated. Power plant systems are calculated for 7000 hr/yr and HDS systems for 330 days/yr (7920 hrs) to arrive at mills/kWh. The TVA study assumes that all newly installed pollution control systems will be operated 7000 hrs/yr for the first two years, so that this is a valid basis for comparison. Limestone is assumed to be available to CAFB and limestone-scrubbing systems at $4/Mg (ton), the cost assumed by TVA. This may be too low, but not enough is yet known about stone requirements to state this with any certainty. The limestone scrubbing units are charged with 25-year ponds for waste slurry containment. It has been assumed that the dry sulfated CAFB waste stone will be removed at no cost to the process.

Labor and laboratory costs have been assumed to remain constant from 50 to 500 MW size units. This may not be strictly accurate, but these costs are such a small part of the total that such an assumption does not significantly affect the results of the comparison. Utility costs have been assumed to be a direct function of plant capacity, a simplifying assumption that also will not affect comparative results. Plant overhead is taken as 20 percent of labor and utility costs, and labor overhead is taken as 10 percent of direct labor and supervision at an average hourly rate of $8.

Operating costs in mills/kWh are plotted against capacity in MW in Figure 3. It is evident from this chart that:

- The present CAFB design is not cost competitive with the limestone scrubbing system design and cost projection and, therefore, must be salable on the basis of
  - Utilization of fuels which cannot be economically used in alternative processes
  - No sludge for disposal
  - Lower water requirements
  - Reliability
- HDS is not competitive, even though operating at a higher load factor, unless a larger than 25,000 bbl/d unit is built to supply more than three 200 MW boilers, or larger than a 35,000 bbl/d unit is built to supply two 500 MW boilers. Thus, HDS requires more immediate commitment of capital than would a single CAFB unit.
- Above about 100 MW, CAFB is lower in cost than a stack-scrubbing retrofit system producing sulfuric acid from the recovered sulfur values.

Catalyst costs for HDS should be noted. These were based on the processing of a West Texas sour vacuum resid having the following properties:

• Boiling range	565°C (1050°F)+
• Sulfur content	5.1 wt %
• API gravity	12.7°
• Conradson carbon number	20
• (V+Ni) content	130 ppm

A Kuwait atmospheric resid having approximately half of the vanadium and nickel content of the West Texas resid also had half of the catalyst costs (6¢/bbl versus 12¢/bbl). Very high heavy metal content resids cannot be economically processed in HDS units. This can be shown if one assumes that catalyst consumption and, thus, catalyst cost are roughly proportional to heavy metal content. The HDS catalyst cost for an Amuay vacuum resid tested in the batch unit at Esso would then be over four times the catalyst cost for West Texas resid, or about 50¢/bbl based on 536 ppm of vanadium and nickel. Furthermore, the Venezuelan Bachaquero vacuum resid would require about $1 of catalyst per bbl of resid, based on 1040 ppm of vanadium and nickle. The CAFB is expected to utilize these resids without cost penalty, but the increased catalyst costs noted above would add 0.71 and 1.65 mills, respectively, to the HDS system costs. These additions to HDS costs would be sufficient to make the largest practical size vacuum resid HDS unit uncompetitive, even though base load operated, compared with a 200 MW 17 percent air/fuel ratio CAFB unit.

The ability of CAFB to handle very heavy high metal content resids must be confirmed by additional tests on the Esso pilot plant. Such an ability could be important, not only in the utilization of petroleum-derived resids but also in association with the production of large quantities of fuel liquids from such sources as the Alberta tar sands, the Orinoco tar belt, the Colorado/Green River oil shale deposits, and highly-volatile, high-sulfur Illinois and western Kentucky coal deposits. Such processing facilities could produce wide boiling range materials generally characterized as "syncrudes" and suitable as feedstocks for existing or new two-stage refinery distillation systems. The fraction of these syncrudes boiling above 565°C (1050°F) would be produced as vacuum bottoms and would also be suitable for 100 percent conversion into clean, noncorrosive fuel gas by CAFB processing. Trace metal contents in such syncrude-derived vacuum bottoms may make conventional coking and combustion of the coke so derived either expensive or physically difficult. For example, ash fusion characteristics of the coke produced from Atabasca tar sands created severe problems in the boilers used to burn that coke. Such problems would be eliminated by the use of CAFB technology. To quote from a recent technical paper by Dr. Kett of Exxon's Florham Park R&D Center, "In the future, progressively more and more of the crude barrel will need to be converted into lower boiling products ..." and the residues from conversion of such lower boiling products, whether from heavier crudes or from syncrudes, can be effectively turned into fuel gas useful in an environmentally acceptable manner by existing power generating boilers.

V. PRELIMINARY DESIGN AND COST ESTIMATE

BACKGROUND

The first phase of the demonstration plant program is to prepare a preliminary design and cost estimate. This phase of the program has included four major tasks. The tasks, participants, and schedule are summarized in Table 10. NEES is the cooperating utility for the demonstration plant program.[2] NEES personnel and the staff at the Narragansett Electric Company, Manchester Street Station provided plant design data and information on plant load requirements, assisted in the selection of the architect-engineer, and participated in the selection of process options and the evaluation of the preliminary design work.

The design effort was based on the experimental data from Esso on its batch units and 1 MW continuous pilot plant, the spent stone processing data from Westinghouse (Appendices K-O), the conceptual design and assessment by Westinghouse,[1,2] and the operating and environmental constraints established by NEES and EPA. EPA has provided general guidance and funding for the Phase I effort. The recommendations developed from this preliminary design and cost estimate will be reviewed by the project participants.

PRELIMINARY DESIGN SCOPE

Three steps were carried out to select the design basis for the CAFB demonstration plant:

- An identification of the process requirements, operating parameters, and design and process options (Appendices J and K)

Table 10

PRELIMINARY DESIGN AND COST ESTIMATE

Task	Participants[a]	Schedule
1. Selection of the architect-engineer and contract negotiation	Westinghouse NEES Esso EPA	January 18, 1973-August 1973
2. Completion of initial design study and cost estimate	Westinghouse SWEC Esso representative on-site consultant Allied Chemical Corporation NEES Esso EPA	September 6, 1973-March 19, 1974
3. Evaluation of the initial design study, selection of the design basis recommended for the demonstration plant, contract negotiation for preliminary design	Westinghouse NEES Esso EPA	March 1974-July 1974
4. Preparation of the preliminary design and cost estimate	Westinghouse SWEC Combustion Engineering NEES Esso EPA	August 6, 1974-December 9, 1974

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[a] Westinghouse served as prime contractor to EPA. NEES is the cooperating utility. Esso has developed the process and is carrying out pilot-scale tests under contract to EPA. SWEC was a subcontractor to Westinghouse to perform engineering services. Allied Chemical Corporation provided design and economic data on sulfur recovery. Combustion Engineering provided an evaluation of boiler performance.

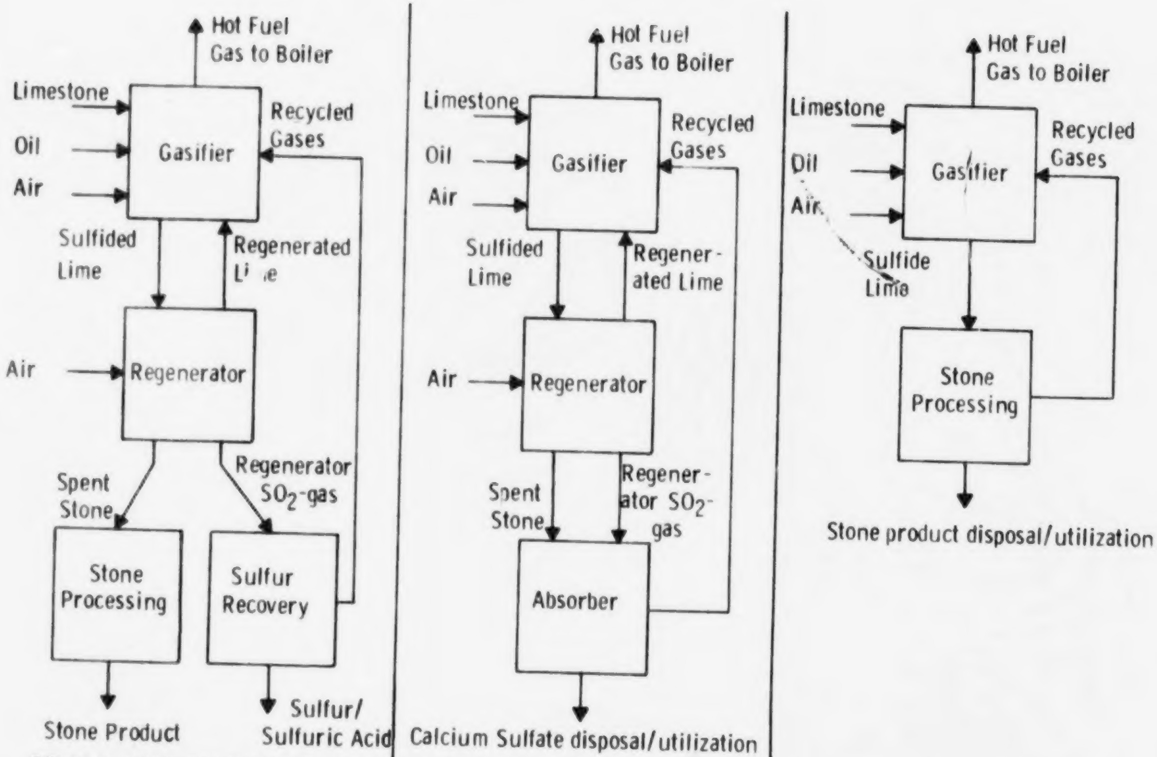
- An initial design study to identify the critical cost and performance factors (Appendix I)
- An assessment of the process options and the initial design study to select the design basis for a demonstration plant which would meet the goals of the program (Appendices J and K).

General Process Options

The general process consists of six processing systems:

- The reaction system (gasification and regeneration)
- The stone processing system
- The sulfur recovery system
- The limestone handling system
- The fuel oil handling system
- The waste/by-product stone handling system.

The general function of the reaction system is to generate a low heating-value fuel gas for utilization by the boiler. The stone processing system converts the spent sorbent into a form suitable for disposal or by-product utilization. The sulfur recovery system produces elemental sulfur from the sulfur captured in the gasifier. The remaining three systems store and carry the raw materials (residual oil or other liquid fuel and limestone) and the waste or by-product materials which are utilized in the process. The three major processing systems can be arranged in a number of ways to generate alternative process schemes. Figure 4 illustrates the general process options. The general regenerative process requires the utilization of stone processing system and a sulfur recovery system arranged in parallel. A number of processes can be utilized for both of these systems. The general once-through process scheme eliminates the regeneration function from the reaction system and requires a stone processing system which produces an environmentally suitable solid product utilizing all of the sulfur removed from the fuel oil. Again, many options exist for this


Hot Fuel Gas to Boiler
Limestone
Oil
Air
Gasifier
Recycled Gases
Sulfided Lime
Regenerated Lime
Air
Regenerator
Spent Stone
Regenerator SO2-gas
Stone Processing
Sulfur Recovery
Stone Product
Sulfur/ Sulfuric Acid
Hot Fuel Gas to Boiler
Limestone
Oil
Air
Gasifier
Recycled Gases
Sulfided Lime
Regener-ated Lime
Air
Regenerator
Spent Stone
Regener-ator SO2-gas
Absorber
Calcium Sulfate disposal/utilization
Hot Fuel Gas to Boiler
Limestone
Oil
Air
Gasifier
Recycled Gases
Sulfide Lime
Stone Processing
Stone product disposal/utilization

GENERAL REGENERATIVE PROCESS

DEMONSTRATION PLANT PROCESS

ONCE-THROUGH PROCESS

Figure 4—Oil gasification processing concepts

stone processing system. A third general processing option is shown in the figure which combines the regenerative and once-through features to eliminate the sulfur recovery system and generate a high sulfur-content solid product.

Numerous process and design options have been identified for the reaction system (Appendix J). These deal with areas such as the technology basis, the market (new boiler versus retrofit, utility versus industrial boiler), advanced concepts yet untested, equipment and operating options. A variety of stone processing options are discussed in Appendices K and O. These include dry sulfation (Appendix L), slurry carbonation (Appendix M), acid sulfation (Appendix N), dead-burning, and dry oxidation followed by dry recarbonation (Appendix O). Alternative processes for sulfur recovery have also been investigated (Appendix P). Considerations such as sulfur recovery versus sulfuric acid production, and existing sulfur recovery processes versus developing technology, have been made.

Initial Design Study

An initial design of the 50 MW CAFB demonstration plant was carried out by SWEC under contract to Westinghouse Research Laboratories. The basis for the initial design study was the utilization of conventional catalytic cracking technology developed in the petroleum industry for the solids transport systems, the specification of equipment to provide extensive system flexibility and reliability, and the inclusion of sulfur recovery from the spent stone. The study identified the critical design parameters and cost items in the plant. These data provided a basis for evaluation and identification of alternative design and operating parameters to reduce the plant cost while maintaining system flexibility and operability.

The critical factors assumed in the design basis for the initial design study are summarized for each of the processing systems.

Reaction System

The initial design basis included items relating specifically to process flexibility and reliability:

- Once-through operation capability in addition to regenerative operation
- 2.44 m (8 ft) bed depth capability
- About 50 percent excess capacity for stack-gas recycle
- Extremely high pressure drops assumed in air and stack-gas circuits: 96.5 kPa (14.0 psi) for control, line losses, bed losses, and distributor losses and equipment design
- Low gasifier plenum start-up temperature and lime storage system for start-up material
- Valves in the fuel-gas lines to isolate the fuel lines for burnout during operation, fuel-gas lines based on 18.3 m/s (60 ft/sec) velocity
- A solids transport system to circulate solids between the gasifier and regenerator based on known catalytic cracker dense phase pneumatic transport technology. Required tall standlegs and relatively high-pressure transport gas
- High elevation of gasifier and regenerator vessels and cyclone fines recycle to reactors by dense phase standleg
- Stainless steel distributor support plate on gasifier
- A stone quench drum used to cool waste solids before they enter the stone processing system.

Sulfur Recovery System

A commercial process to recover elemental sulfur from the regenerator product gas (8 volume percent SO_2) was specified. The process (Allied Chemical) is described in Appendix P.

Stone Processing System

The stone processing system selected for the initial design study was a slurry carbonation process which removed carbon dioxide (CO_2) from the plant stack gas to contact the regenerator waste stone and produce a dry calcium carbonate product. The hydrogen sulfide (H_2S) produced in the process is recycled to the sulfur recovery system.

Limestone Handling and Feeding System

The design basis for the limestone handling and feeding system used in the initial study is four-times the stoichiometric rate.

Waste Stone Handling System

The design basis for the waste stone handling system is a four-times-stoichiometric limestone feed rate.

Table 11 lists the total installed costs of the six process systems in the initial study demonstration plant in end-of-1974 dollars. It is evident from the table that the reaction and sulfur recovery systems represent the dominant costs in the initial study demonstration plant, or 77 percent of the total plant cost. The major items in the reaction system cost appear to be gas compression equipment at 30 percent of the reaction system cost. Detailed costs are presented in Appendix I.

The initial design study has been used to guide the selection of a CAFB demonstration plant design basis. Specifically it has indicated that:

- The commercial sulfur recovery process applied in the study results in a significant cost penalty for this application.
- Conventional technology utilized by the petroleum industry for solids circulation is considered excessively expensive in this application. Techniques developed in the pilot-plant operation will result in a more economical demonstration plant.

Table 11

SUMMARY OF MAJOR SYSTEM COSTS IN
THE INITIAL STUDY DEMONSTRATION PLANT

System	Total installed cost (1000 $)
Reaction	4,511.00
Stone Processing (slurry carbonation)	1,102.00
Sulfur Recovery	3,000.00
Waste Stone Handling	358.00
Fuel Oil Handling	825.00
Limestone Handling	174.00
TOTAL PLANT	9,970.00

- The critical cost items in the reaction system are the gas-compression components and the fuel-gas handling components, and these may be greatly reduced in cost by proper design.
- A stone processing system should be utilized in the demonstration plant which permits the elimination of a sulfur recovery system (Appendix K).

Selection of the Preliminary Design Basis

The selection of the design basis for the demonstration plant from the identified process and design options and the results of the initial design study rests on a number of general factors:

- Maximum demonstration plant operability based on pilot-plant results
- Flexibility to gain maximum information from the plant
- Applicability of the demonstration plant to commercial plant considerations
- Environmental performance and goals
- Economics of the demonstration plant and economic targets for a commercial plant.

These factors have been utilized in the design basis selection, in addition to the requirement that the demonstration plant function as a commercial power plant upon completion of the demonstration plant program. This selection process is described in Appendix J and results in the adoption of the conclusion drawn from the initial design study.

PRELIMINARY DESIGN

Design Basis

Some general factors assumed in the design and cost estimate for the demonstration plant are listed below:

- Retrofit to New England Power Service Company No. 12 boiler, Manchester Street Station, Narragansett Electric Company, Providence, R. I.

- Cost of removal of No. 10 and No. 11 boilers to provide space is not included, and cost of refractory-lined 1067 mm (42 in) piping for hot fuel gas is based on four burners located together in near side of No. 12 boiler. Tangential firing will require at least 50 percent more 1067 mm (42 in) piping.
- Boiler house, overhead coal bunker, ash silos, railroad siding, rail car spotting equipment, and oil barge unloading facilities are all existing and usable.
- Steam, boiler feedwater, service cooling water, electrical power, plant air, instrument air, and low-sulfur fuel oil are all available by connection to existing supplies at no cost.
- Continuous firing of present normal fuel at a 5 percent minimum of boiler capacity; 50 MW power requires 191,000 kg/hr (422,000 lb/hr) steam. The hot fuel gas is estimated to raise 195,000 kg/hr (430,000 lb/hr). The boiler capacity is 204,000 kg/hr (450,000 lb/hr).
- Fuel oil feed weight-off has sufficiently fast response to boiler fuel rejection signals so there is no need for hot fuel-gas line to have cut-off and relief facilities: if the latter were required, significant cost and major practical problems would be entailed.
- The design parameters and effectiveness of absorber reaction at 94 percent minimum efficiency (not more than 6 percent of sulfur dioxide and of calcium oxide unreacted) will be tested and proved in the pilot plant.
- Complete reaction of residual calcium sulfide in regenerated stone occurs in the absorber (so that waste stone is unreactive toward natural waters) and will be tested and proved in the pilot plant.
- Suitable limestone supply is 483 km (300 miles) by rail. Suitable waste solids disposal is 48 km (30 miles) by truck.

- Horizontal pulsed flow for controlled stone circulation will be tested and proved in a full-scale cold flow model.
- Air jet pulverizing of regenerated stone for feed to absorber will be tested and proved in manufacturer's test equipment.
- The 50:1 scale-up of processes proved in the Esso pilot plant is feasible and appropriate.

A detailed summary of the design basis and assumptions is given in Appendix D. Important factors in the design specification are listed.

- Plant size - 46 MW
- Fuel oil rate - 12,700 kg/hr (28,000 lb/hr)
- Plant sulfur removal efficiency - 90%
- Plant turndown ratio - 4:1
- Fuel oil properties - Table D-1
- Limestone properties - Table D-1; a final selection of limestone for the plant has not yet been made. Laboratory tests and pilot-plant tests are being carried out in order to make the selection (Appendix Q).
- Gasification process concept selected - regenerative process with dry sulfation of spent stone
- Reaction section conditions

  Gasifier conditions/design

    Operating temperature - 880°C (1616°F)

    Pressure - boiler pressure plus pressure losses through cyclones, gas ducts, and burners 17.9 kPa (2.6 psi)

    Air/fuel ratio - 20% of stoichiometric maximum (includes oxygen in flue gas recycle)

    Superficial fluidization velocity - 1.83 m/s (6 ft/sec)

    Bed depth (maximum) - 1.22 m (4 ft)

    Sulfur removal efficiency - to yield 90 percent plant efficiency

    Limestone makeup rate - 0.5 to 1.5 times stoichiometric

    Number of gasifier modules - 1

    Gasifier temperature control - flue gas recycle with provision for steam and water injection

Burnout - provision for continuous steam/air injection into cyclone inlets/or burnout with flue gas during shutdown
Turndown ratio - 4/1
Turndown method - control of air/fuel ratio with minimum velocity reduction
Number of fuel-gas ducts - 4 (one per burner)
Elutriation rate of sorbent from gasifier - 3,175 kg/hr (7,000 lb/hr)

Regenerator conditions/design
  Operating temperature - 1080°C (1976°F)
  Pressure - 0 to 2.5 Pa (0 to 10 in. H_2O) less than the gasifier pressure
  Air rate - 0.1% oxygen in product gas
  Superficial velocity - 1.83 m/s (6 ft/sec)
  Bed depth - 1.22 m (4 ft)
  Number of modules - 1
  Temperature control - by stone circulation rate emergency inert gas quench for temperature overshoot

Solids circulation system
  Esso dense phase pulsed flow technique

Gasifier cyclones and fines circulation system
  Four cyclones on gasifier set at minimum elevation
  Cyclone fines returned to regenerator by commercial plug flow technique

Air circuits/flue gas circuit pressure drops
  Minimize losses by proper design of blower control, gas lines, and so on
  Single blower for air and stack gas to gasifier

The reaction system in the demonstration plant is designed on the basis of the best existing pilot-plant data, but the demonstration plant is also designed with the flexibility of an experimental installation to reflect the uncertainties involved in the scale-up of the pilot-plant information, the integration of the process with a full-scale power plant, the turndown performance of the plant, and operating procedures; and also to allow the testing of many operating and design options. The capacity and complexity of all the components in the reaction system are increased by this requirement for flexibility.

- Dry sulfation system
  Pulverizer
    Spent stone reduced in size to about 75 micron average diameter while hot in commercial jet pulverizer (Jet-O-Mizer, Fluid Energy Processing and Equipment Company, Hatfield, Pa.)
  Absorber
    Bed temperature - 870°C (1598°F)
    Fluidization velocity - 0.122 m/s (0.40 ft/sec)
    SO_2 absorption efficiency - 90 percent

Energy and Material Balances

Common assumptions have been made for generating energy and material balances having sufficient accuracy for a preliminary design task. These include:

- No heat losses
- Steady state with no accumulations
- Air to blowers is at 25°C (77°F) and bone dry
- Purge and transport gas streams have been neglected
- The pulsed nature of solids transport systems have been neglected by assuming steady flow behavior

When possible, data and correlations generated by the Esso experimental programs have been utilized to carry out balances -- for example, for fuel-gas compositions and gasifier heat generation rates. Fines carried over in the gas streams have been included in the balances because they are an important operating and environmental consideration. Material and energy balances have been carried out for three cases:

- The base case condition at full capacity
- A case with plant turndown to 25 percent of capacity
- A case with a 17 percent of stoichiometric air/fuel ratio as compared with the base value of 20 percent of stoichiometric.

The balances are summarized in Figures 5 and 6.

Demonstration Plant Design

The Manchester Street Station, boiler No. 12, is a 46 MW, balanced draft, CE unit producing 204,120 kg (450,000 lb) steam/hr at 510°C (950°F) and 83 bars (1200 psig). The boiler is designed for both coal- and oil-firing. The plant site is described in Appendix C.

The demonstration plant is a facility which will demonstrate system operability, environmental performance, boiler performance, and process retrofit capability. It will also provide a basis for establishing commercial plant economics. The demonstration plant has been designed with sufficient flexibility to enable the exploration and determination of optimum operating conditions and operating procedures.

The demonstration plant consists of five major processing systems:

- Reaction
- Stone processing
- Limestone handling
- Fuel-oil handling
- Waste/by-product stone handling

The general function of the reaction system is to generate a low-heating-value fuel gas for utilization by the boiler. The stone processing system employs a dry sulfation process to convert the spent sorbent into a form suitable for disposal or by-product utilization. The remaining three systems store and carry the raw materials (residual oil and limestone) and the waste or by-product materials (sulfated lime) which are utilized in the process. A detailed flow diagram of the reaction system and the dry sulfation system is shown in Figure 7. Numbered streams in the figure correspond to the numbered streams in the material and energy balance figures.

Reaction System

The reaction system consists of

- A gasifier
- A regenerator
- Solids transport components
- Particulate removal components
- Air and stack-gas compression components
- Fuel-gas handling and combustion components
- Heat exchange components
- A purge and transport gas (PTG) system.

Residual fuel oil is fed to the gasifier at 12,700 kg/hr (28,000 lb/hr) along with limestone at 1,070 kg/hr (2,355 lb/hr), air at 32,100 kg/hr (70,604 lb/hr) and stack gas for temperature control at 30,800 kg/hr (67,020 lb/hr). Oil cracking, partial oxidation, and hydrogen sulfide (H_2S) absorption proceed at 880°C (1616°F) in the gasifier to generate 76,000 kg/hr (167,175 lb/hr) of low heating-value fuel gas. The hot fuel gas flows to four boiler burners through four separate refractory-lined pipes where combustion is completed for steam generation. The sulfided lime ($\sim$ 6 wt % calcium sulfide [CaS]) is transported pneumatically to the regenerator vessel at 31,400 kg/hr (69,419 lb/hr). The calcium sulfide in the stone and the carbon ($\sim$ 0.3 wt % $CH_{1/2}$) are reacted with preheated air at a temperature of 1075°C (1970°F) to generate calcium oxide (CaO), calcium sulfate, sulfur dioxide (SO_2) (8 mole %), and water vapor and carbon dioxide. Regenerated sorbent is recirculated to the gasifier at 32,400 kg/hr (71,615 lb/hr). The sulfur dioxide gas at 3,980 kg/hr (8,778 lb/hr)

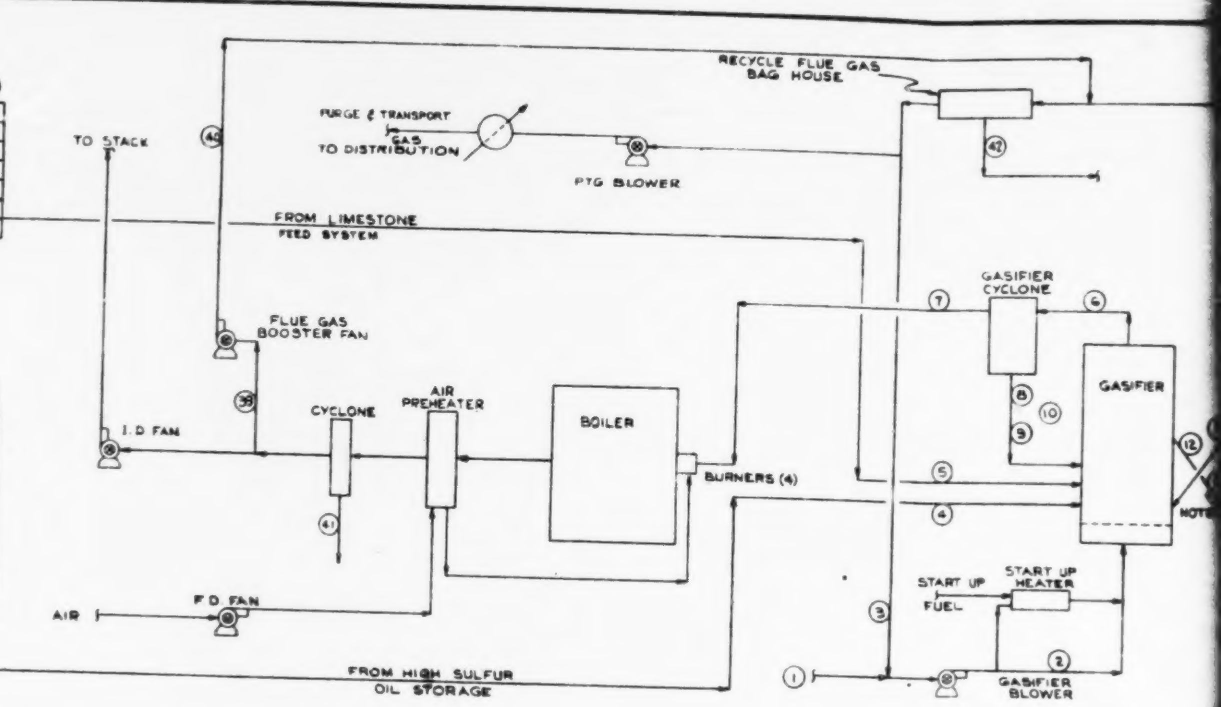
BASE CASE LIMESTONE ⑤

S	2.8
CaO	1274.0
CO_2	1036.2
INERTS	41.9
TOTAL	2354.9
TEMP.	77
PRESS	22.0

HIGH SULFUR FUEL OIL ④

	BASE CASE	TURNDOWN CASE
C	24052.0	6013.0
H	3080.0	770.0
O	28.0	7.0
N	112.0	28.0
S	728.0	182.0
TOTAL	28000.0	7000.0
V	12.4	3.1
Na	0.8	0.2
Ni	1.6	0.4
TEMP.	200	200
PRESS	25.0	25.0

NOTE 2


TO STACK
㊵
PURGE & TRANSPORT GAS TO DISTRIBUTION
PTG BLOWER
RECYCLE FLUE GAS BAG HOUSE
㊷
FROM LIMESTONE FEED SYSTEM
FLUE GAS BOOSTER FAN
㊴
I.D FAN
CYCLONE
AIR PREHEATER
BOILER
BURNERS (4)
㊶
AIR
F.D FAN
GASIFIER CYCLONE
GASIFIER
⑦
⑥
⑧
⑨
⑩
⑤
④
⑫
NOTE
START UP FUEL
START UP HEATER
③
①
②
GASIFIER BLOWER
FROM HIGH SULFUR OIL STORAGE

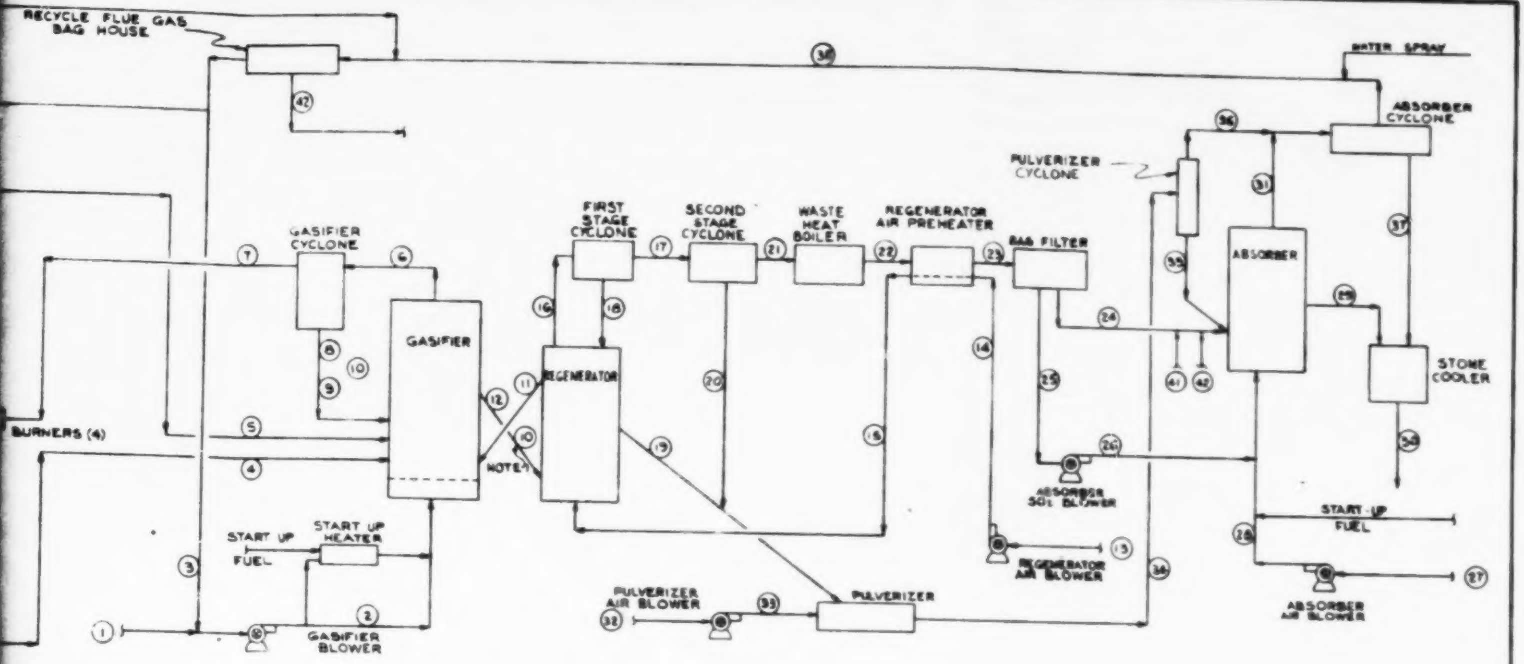
BASE CASE MATERIAL BALANCE - NORMAL OPERATION - QUANTITIES IN (LB/HR)

STREAM	①	②	③	⑥	⑦	⑧	⑨	⑩	⑪	⑫
COMPONENT	GASIFIER AIR	GAS FR AIR RECYCLE	TOTAL RE CYCLE GAS	FUEL GAS	FUEL GAS	GASIFIER FINES	FINES TO INLET	FINES TO OUTLET	REGEN. STONE	SULFIDE STONE
N_2	54223.9	104676.3	50452.4	104788.3	104788.3					
O_2	16380.1	17898.8	1518.7							
CO				12651.5	12651.5					
CO_2		12086.3	12086.3	27316.6	27316.6					
H_2O		5802.3	5802.3	7027.5	7027.5					
H_2				750.0	750.0					
CH_4				3500.0	3500.0					
C_2H_4				5600.0	5600.0					
$>C_2$				7461.3	7461.3					
H_2S				77.1	77.1					
SO_2		160.3	160.3							
CaO						6141.6	3070.8	3070.8	65288.3	61988.6
CaS						404.0	202.0	202.0	2184.0	4077.5
$CaSO_4$						84.8	42.4	42.4	1762.9	855.8
$CH_{1/2}$						19.6	9.8	9.8		198.0
INERTS						227.0	113.5	113.5	2379.6	2296.0
TOTAL	70604.0	140624.0	70020.0	169172.3	169172.3	6877.0	3438.5	3438.5	71614.8	69418.7
FINES				7239	362					
TEMP °F	77	325	350	1616	1616	1616	1616	1616	1970	1616
PRESS PSIA	14.6	23.3	15.0	17.3	16.8	16.8	19.0	19.0	18.8	18.5
PTG.							30.0	30.0	580.	580.

STREAM	⑬	⑭	⑮	⑯	⑰	⑱	⑲	⑳	㉑	㉒	㉓	㉔	㉕	㉖	㉗	㉘	㉙	㉚
COMPONENT	REGEN. AIR	REGEN AIR	REGEN AIR	SO_2 GAS	SO_2 GAS	1ST STAGE FINES	REGEN WASTE	2ND STAGE FINES	SO_2 GAS	SO_2 GAS	SO_2 GAS	SO_2 GAS FINES	SO_2 GAS	SO_2 GAS	ABSORBER AIR	ABSORBER AIR	SULFATED STONE	SOLID WASTE
N_2	6544.6	6544.6	6544.6	6544.6	6544.6				6544.6	6544.6	6544.6		6544.6	6544.6	2530.3	2530.3		
O_2	1977.0	1977.0	1977.0	9.7	9.7				9.7	9.7	9.7		9.7	9.7	764.3	764.3		
CO																		
CO_2				731.2	731.2				731.2	731.2	731.2		731.2	731.2				
H_2O				74.6	74.6				74.6	74.6	74.6		74.6	74.6				
H_2																		
CH_4																		
C_2H_4																		
$>C_2$																		
H_2S																		
SO_2				1417.6	1417.6				1417.6	1417.6	1417.6		1417.6	1417.6				
CaO						94.3	597.3	282.9				18.4					37.1	84.0
CaS						3.2	20.0	9.5				0.6						1.5
$CaSO_4$						2.5	16.1	7.6				0.5					25056	2793.5
$CH_{1/2}$																		
INERTS						3.4	21.8	10.2				0.7					37.6	40.6
TOTAL	8521.6	8521.6	8521.6	8777.7	8777.7	103.4	655.2	310.2	8777.7	8777.7	8777.7	20.2	8777.7	8777.7	3294.6	3294.6	2580.9	2919.6
FINES				454.3	330.9				20.7	20.7	20.7		0.5	0.5				
TEMP °F	77	158	452	1970	1970	1970	1970	1970	970	600	300	300	300	409	77	147	1600	200
PRESS PSIA	14.6	24	23	17.0	16.6	16.6	17.0	15.8	15.8	14.8	14.4	14.0	14.0	22.5	14.6	22.5	16.5	14.7
PTG.						0.2	6.	0.6				0.1					9.2	

NOTE:

1. STREAM 10 IS INJECTED NEAR GASIFIER OUTLET TO MIX WITH STREAM 12 AND PASS TO THE REGENERATOR. STREAM 8 IS DIVIDED INTO 9 & 10 - 50% IN EACH STREAM
2. APPROX. 100% OF V, 36% OF Na AND 75% OF Ni ARE RETAINED IN WASTE STONE STREAM 30


RECYCLE FLUE GAS BAG HOUSE
WATER SPRAY
ABSORBER CYCLONE
PULVERIZER CYCLONE
GASIFIER CYCLONE
FIRST STAGE CYCLONE
SECOND STAGE CYCLONE
WASTE HEAT BOILER
REGENERATOR AIR PREHEATER
BAG FILTER
ABSORBER
STONE COOLER
GASIFIER
REGENERATOR
NOTE 1
BURNERS (4)
START UP FUEL
START UP HEATER
GASIFIER BLOWER
PULVERIZER AIR BLOWER
PULVERIZER
ABSORBER SO2 BLOWER
REGENERATOR AIR BLOWER
START-UP FUEL
ABSORBER AIR BLOWER
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42

(21) SO_2 GAS	(22) SO_2 GAS	(23) SO_2 GAS	(24) SO_2 GAS FINES	(25) SO_2 GAS	(26) SO_2 GAS	(27) ABSORBER AIR	(28) ABSORBER AIR	(29) SULFATED STONE	(30) SOLID WASTE	(31) ABSORBER GAS	(32) PULVERIZER AIR	(33) PULVERIZER AIR	(34) PULVERIZER PROD.	(35) ABSORBER FEED	(36) PULVERIZER AIR	(37) ABSORBER FINES	(38) ABSORBER RECYCLE	(39) STACK GAS RECYCLE	(40) STACK GAS RECYCLE	(41) BOILER FINES	(42) RECYCLE FINES	STACK GAS	BOILER AIR
6544.6	6544.6	6544.6		6544.6	6544.6	2530.3	2530.3			8608.6	1941.0	1941.0	1941.0		1941.0		10549.6	39902.8	39902.8			316360.6	249008.9
9.7	9.7	9.7		9.7	9.7	764.3	764.3			292.7	586.3	586.3	586.3		586.3		879.0	639.7	639.7			5026.7	75221.4
731.2	731.2	731.2		731.2	731.2					731.2							731.2	11355.1	11355.1			84227.2	
74.6	74.6	74.6		74.6	74.6					74.7							2274.7	3527.6	3527.6			27719.5	
1417.6	1417.6	1417.6		1417.6	1417.6					141.2							141.2	18.5	18.5			143.6	
			18.4					37.1	84.0				880.2	836.2		46.0				288.8	7.2		
			0.6						1.5				29.5	28.0		1.5					0		
			0.5					2505.6	2793.5				25.6	22.5		287.9				3.8	9.0		
			0.7					37.6	40.6				32.0	30.4		5.8				10.2	0.4		
8777.7	8777.7	8777.7	20.2	8777.7	8777.7	3294.6	3294.6	2580.5	2919.6	9849.0	2527.3	2527.3	3452.8	917.1	2527.3	341.1	14576.3	55443.7	55443.7	302.8	16.6	436479.6	324230.3
20.7	20.7	20.7		0.5	0.5					304.4					48.5		10.6	6.0	6.0			47.2	
1970	600	300	300	300	409	77	147	1600	200	1600	77	395	794	794	794	1435	500	300	310	300	500	300	
15.8	14.8	14.4	14.0	14.0	22.5	14.6	22.5	16.5	14.7	16.5	14.6	75	16.8	16.5	16.5	15.8	15.8	14.2	14.9	14.2	14.6		
			0.1					5.2						2		0.8					0.2		

SYSTEM

ISSUE	DATE	DESCRIPTION OF ISSUE	BY	CHK	APP'D
1	[illegible]	GENERAL REVISIONS BY WESTINGHOUSE	HDR		
P1	10/7/74	WITH WESTINGHOUSE APPROVAL	HDR		
P	9.25.74	PRELIMINARY	HDR		

ENGINEERED

DRAWN

CHECKED

APPROVED

CERTIFIED FOR CONSTRUCTION

BY

DATE

STONE & WEBSTER ENGINEERING CORPORATION

CLIENT

WESTINGHOUSE ELECTRIC CORPORATION
FLUID BED OIL GASIFICATION
DEMONSTRATION PLANT
PROVIDENCE, RHODE ISLAND

TITLE

SCHEMATIC FLOW DIAGRAM
AND MATERIAL BALANCE
BASE CASE

Figure 5

ISSUE
1

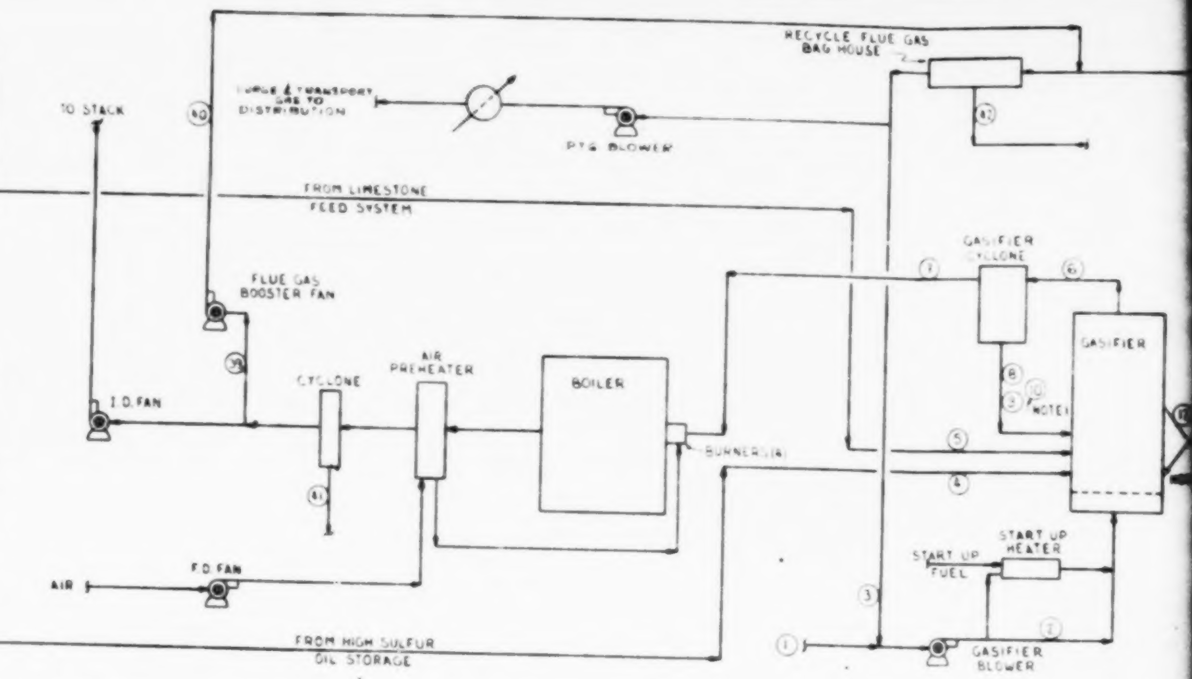
TURNDOWN CASE LIMESTONE ⑤

S	[illegible]
CaO	318.5
CO_2	259.0
INERTS	11.2
TOTAL	589.4
TEMP	77
PRESS	22.0

HIGH SULFUR FUEL OIL ④

	BASE CASE	TURNDOWN CASE
C	24052.0	6013.0
H	3080.0	770.0
O	28.0	7.0
N	112.0	28.0
S	728.0	182.0
TOTAL	28000.0	7000.0
V	12.4	3.1
Na	0.3	.075
Ni	1.1	.275
TEMP	200	200
PRESS	23.0	23.0

NOTE 2



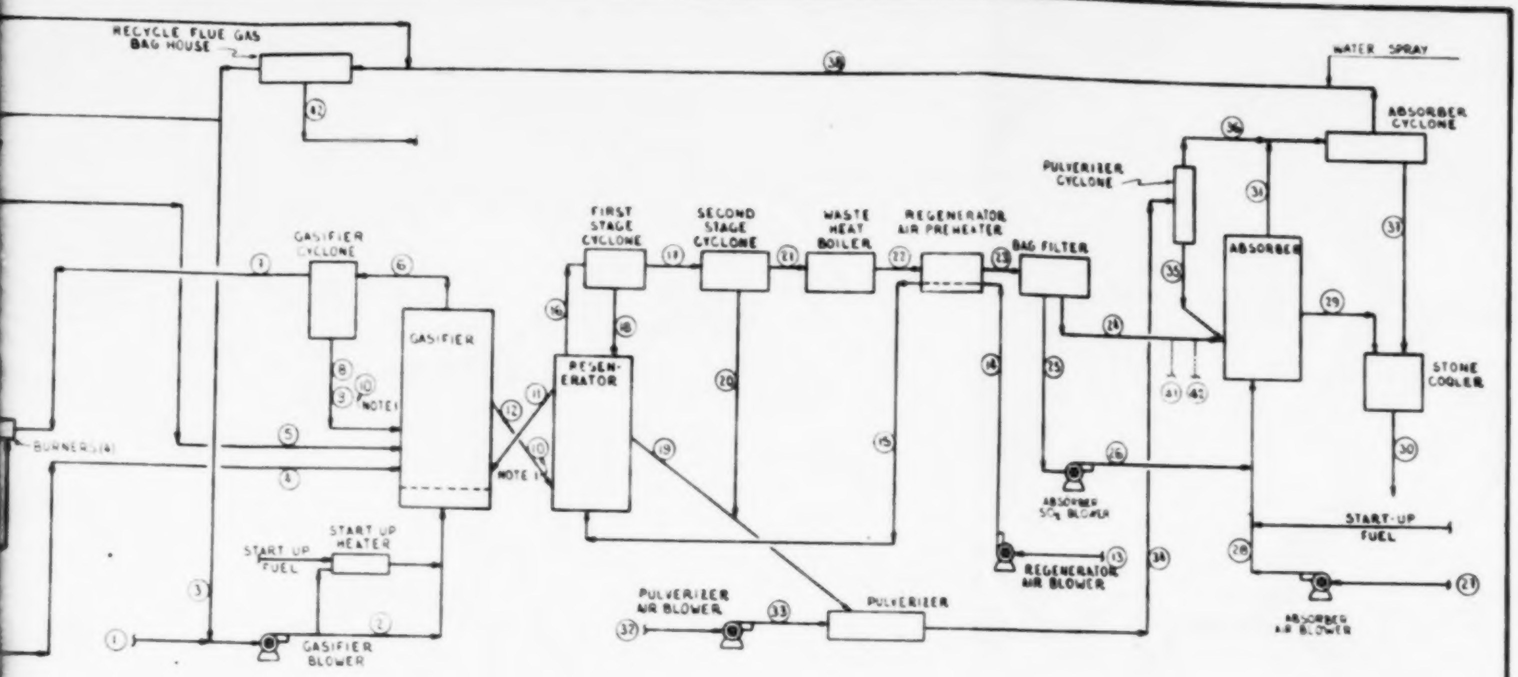
TO STACK
40
PURGE & TRANSPORT GAS TO DISTRIBUTION
PTG BLOWER
RECYCLE FLUE GAS BAG HOUSE
42
FROM LIMESTONE FEED SYSTEM
FLUE GAS BOOSTER FAN
39
I.D. FAN
CYCLONE
41
AIR PREHEATER
BOILER
BURNERS(4)
GASIFIER CYCLONE
7
6
GASIFIER
8
9
10
NOTE 1
5
4
11
START UP FUEL
START UP HEATER
3
1
2
GASIFIER BLOWER
AIR
F.D. FAN
FROM HIGH SULFUR OIL STORAGE

TURNDOWN CASE MATERIAL BALANCE - QUANTITIES IN LB/HR (NOTE 3)

STREAM	①	②	③	⑥	⑦	⑪	⑫	⑬	⑭	⑮	⑯	⑰	⑱	⑲	⑳	㉑	㉒	㉓	㉔	㉕	㉖	㉗	㉘	㉙	㉚	㉛	㉞	㉟
COMPONENT	GASIFIER AIR	GASIFIER AIR RECYCLE	TOTAL RECYCLE GAS	FUEL GAS	FUEL GAS	REGEN STONE	SULFIDE STONE	REGEN AIR	REGEN AIR	REGEN AIR	SO_2 GAS	SO_2 GAS	1st STAGE FINES	REGEN WASTE	2nd STAGE FINES	SO_2 GAS	SO_2 GAS	SO_2 GAS	SO_2 GAS FINES	SO_2 GAS	SO_2 GAS	ABSORBER AIR	ABSORBER AIR FUEL	SULFATED STONE	SOLID WASTE	ABSORBER GAS	PULVER. AIR PROD	ABSORBER FEED
N_2	7399.4	56050.1	48650.6	56098.2	56098.2			3450.9	3450.9	3450.9	3451.0	3451.0				3451.0	3451.0	3451.0		3451.0	3451.0	2441.9	2441.9			5826.3	1941.0	
O_2	2239.3	2234.2	4958.9					1042.5	1042.5	1042.5	587.6	587.6				587.6	587.6	587.6		586.4	587.6	738.3	738.3			816.6	586.3	
CO				5249.8	5249.8																							
CO_2		12239.9	12239.9	19817.1	19817.1																							
H_2O		2332.3	2332.3	2985.1	2985.1																					342.3		
H_2				100.0	100.0																					48.2		
CH_4				1200.0	1200.0																							
C_2H_6				300.0	300.0																							
HCs				1131.9	1131.9																							
H_2S				25.2	25.2																							
SO_2		44.2	44.2																									
CaO						8051.3	8051.2				350.6	350.6				350.6	350.6	350.6		350.6	350.6					35.0		
CaS							388.5						24.6	304.5	65.9				9.2					15.2	30.5		306.5	291.2
$CaSO_4$						163.9	440.6						8		2.2				3									
													3	29.1	1.8				25					698.1	698.6		29.1	27.6
INERTS						298.0	305.2						9	11.1	2.4				35					10.6	11.1		11.1	10.6
FUEL NO.2																							109.3					
TOTAL	9634.3	72390.8	58360.0	85287.3	85287.3	9109.2	9351.5	4493.4	4493.4	4493.4	4389.0	4389.0	21.0	346.7	72.3	4389.0	4389.0	4389.0	[illegible]	4389.0	4389.0	3184.8	3294.6	723.9	741.2	7285.[illegible]	2844.0	329.6
FINES											110	83				10.4	10.4	10.4								200		
TEMP °F	77			1550	1550	1970	1550	77	77	77	1970	1970	1970	1970	1970	1970	600	300	300	300	300			1600	200	1600	602	609
PRESS PSIA	14.6							14.6	14.6	14.6			16.6		15.8				14.0						14.7			
PTG													0.2		0.6				0.1									

NOTE

1. STREAM 10 IS INJECTED NEAR GASIFIER OUTLET TO MIX WITH STREAM 12 AND PASS TO THE REGENERATOR. STREAM 8 IS DIVIDED INTO 9 & 10 - 50% IN EACH STREAM.
2. APPROX. 100% OF V, 96% OF Na, AND 70% OF Ni ARE PICKED UP BY WASTE STONE, STREAM 30.
3. STREAMS NOT SHOWN OTHERWISE ARE SAME AS FOR BASE CASE.


RECYCLE FLUE GAS BAG HOUSE
(38)
(42)
WATER SPRAY
ABSORBER CYCLONE
(36)
PULVERIZER CYCLONE
(34)
(37)
GASIFIER CYCLONE
FIRST STAGE CYCLONE
SECOND STAGE CYCLONE
WASTE HEAT BOILER
REGENERATOR AIR PREHEATER
BAG FILTER
ABSORBER
(1)
(6)
(17)
(21)
(22)
(23)
(35)
(16)
(18)
(24)
(29)
GASIFIER
REGEN-ERATOR
(8)
(9)
(10)
NOTE 1
(11)
(12)
(20)
(14)
(25)
(41)
(42)
STONE COOLER
BURNERS (4)
(5)
(4)
(10)
NOTE 1
(19)
(15)
(26)
(30)
ABSORBER SO2 BLOWER
START-UP FUEL
START UP HEATER
START UP FUEL
(3)
(13)
REGENERATOR AIR BLOWER
(34)
(28)
PULVERIZER AIR BLOWER
PULVERIZER
(27)
ABSORBER AIR BLOWER
(2)
(1)
GASIFIER BLOWER
(32)
(33)

(24) SO2 GAS FINES	(25) SO2 GAS	(26) SO2 GAS	(27) ABSORBER AIR	(28) ABSORBER AIR/FUEL	(29) SULFATED STONE	(30) SOLID WASTE	(31) ABSORBER GAS	(34) PULVERIZER PROD	(35) ABSORBER FEED	(37) ABSORBER FINES	(38) ABSORBER RECYCLE	(39) STACK GAS RECYCLE	(40) STACK GAS RECYCLE	STACK GAS	○
	3451.0	3451.0	2885.9	2445.9			5836.9	1941.0			7837.9	80812.8	80812.8	118532.6	
	5836.6	5836.6	736.3	736.3			866.8	586.3			1453.1	35458	35458	6710.0	
							342.3				342.3	11937.6	11937.6	22689.0	
							166.1				1866.2	484.1	480.1	518.3	
	350.6	350.6					35.0				15.0	9.2		34.8	
9.2					15.2	50.5		306.5	291.2	15.3					
3															
25					638.1	639.6		29.1	27.6	1.5					
.35					10.6	11.1		11.1	10.6	0.5					
				109.8											
10.1	4389.0	4383.0	3184.2	3294.6	715.3	741.3	4285.2	2814.0	329.4	17.3	1156.5	56183.5	56183.5	101144.3	
							200			280	8				
300	300	300			1600	200	1600	603	603	342	500	300	310		
14.0						14.4						14.2	14.9		
0.1															

LOW AIR/FUEL RATIO CASE · QUANTITIES IN LB/HR (NOTE 5)

STREAM COMPONENT	(1) GASIFIER AIR	(2) GASFR AIR RECYCLE	(3) TOTAL AIR CYCLE GAS	(6) FUEL GAS	(7) FUEL GAS	(39) STACK GAS RECYCLE	(40) STACK GAS RECYCLE	STACK GAS
N_2	45055.4	38904.4	38309.0	83516.4	83516.4	27759.4	27759.4	38313.7
O_2	13622.6	15213.9	1591.3			912.3	912.3	8065.6
CO				11058.6	11058.6			
CO_2		8641.5	8641.5	20961.9	20961.9	7880.3	7880.3	
H_2O		4712.9	4712.9	6913.2	6913.2	2448.2	2448.2	
H_2				300.0	300.0			
CH_4				3200.0	3200.0			
C_2H_6				5600.0	5600.0			
$>C_3$				9227.6	9227.6			
H_2S				36.1	36.1			
SO_2		158.3	158.3			12.9	12.9	145.6
CaO								
CaS								
$CaSO_4$								
INERTS								
FUEL NO 2								
TOTAL	58718.0	42,107.4	53,389.4	140,847.8	140,847.8	38,813.1	38,813.1	43941.6
FINES				7239.0	362.0	6.0	6.0	47.2
TEMP °F	77			1665	1665	300	310	300
PRES PSIA	16.6	23.3	15.0	17.3	16.3	16.2	14.9	
PTG								

SYSTEM

ISSUE	DATE	DESCRIPTION OF ISSUE	BY	CHK	APPVD
1	10/21/80	GENERAL REVISIONS BY WESTINGHOUSE	HDR		
P	10/27/80	WITH WESTINGHOUSE APPROVAL	HDR		
P	9/25/80	PRELIMINARY	HDR		

ENGINEERED DRAWN CERTIFIED FOR CONSTRUCTION BY
CHECKED APPROVED DATE

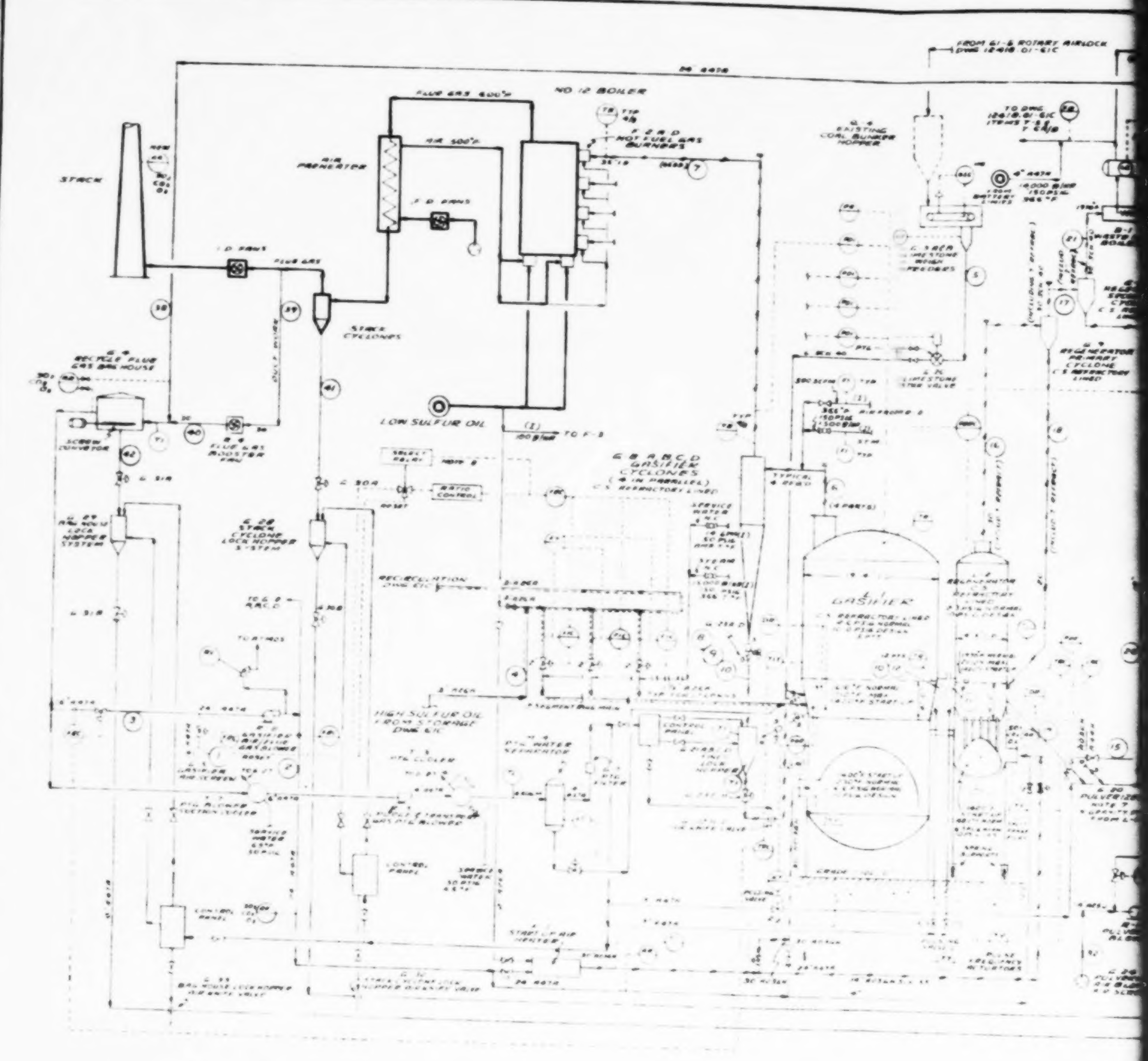
STONE & WEBSTER ENGINEERING CORPORATION

CLIENT
WESTINGHOUSE ELECTRIC CORPORATION
FLUID BED OIL GASIFICATION
DEMONSTRATION PLANT
PROVIDENCE, RHODE ISLAND

TITLE
SCHEMATIC FLOW DIAGRAM
MATERIAL BALANCE
TURNDOWN CASE AND
LOW AIR/FUEL RATIO CASE

ISSUE
1

Figure 6


FROM G1-6 ROTARY AIRLOCK DWG 12418-01-61C
24" A4TA
NO 12 BOILER
FLUE GAS 600°F
AIR 500°F
AIR PREHEATER
F.D. FANS
F 2 & D HOT FUEL GAS BURNERS
STACK
I.D. FANS
FLUE GAS
STACK CYCLONES
G-4 RECYCLE FLUE GAS BAG HOUSE
SCREW CONVEYOR
B-6 FLUE GAS BOOSTER FAN
LOW SULFUR OIL
TO F-3
G-4 EXISTING COAL BUNKER HOPPER
TO DWG 12418-01-61C
4" A4TA
14,000 LB/HR 150 PSIG 366°F
FROM BATTERY LIMITS
G-3 A,B LIMESTONE WEIGH FEEDERS
L-26 LIMESTONE FEEDER VALVE
G-9 REGENERATOR PRIMARY CYCLONE C.S. REFRACTORY LINED
B-1 WASTE HEAT BOILER
G-8 A,B,C,D GASIFIER CYCLONES (4 IN PARALLEL) C.S. REFRACTORY LINED
SELECT RELAY
RATIO CONTROL
SERVICE WATER N.C.
STEAM N.C.
TYPICAL 4 REQD
(4 PARTS)
L-1 GASIFIER C.S. REFRACTORY LINED
L-2 REGENERATOR C.S. REFRACTORY LINED
G-29 BAG HOUSE LOCK HOPPER SYSTEM
G-28 STACK CYCLONE LOCK HOPPER SYSTEM
RECIRCULATION
HIGH SULFUR OIL FROM STORAGE
PTG WATER SEPARATOR
CONTROL PANEL
SERVICE WATER
STARTUP AIR HEATER
GRADE
PULSE FREQUENCY ACTUATORS
G-33 BAG HOUSE LOCK HOPPER AIR KNIFE VALVE
G-20 PULVERIZER

NOTES: 1 HEAVY LINES INDICATE EXISTING EQUIPMENT

2 GRADE ELEVATION = 100'-0

3 —=— DENOTES REFRACTORY LINED C.S. PIPE ([illegible] SPECIAL)

4 (I) = INTERMITTENT

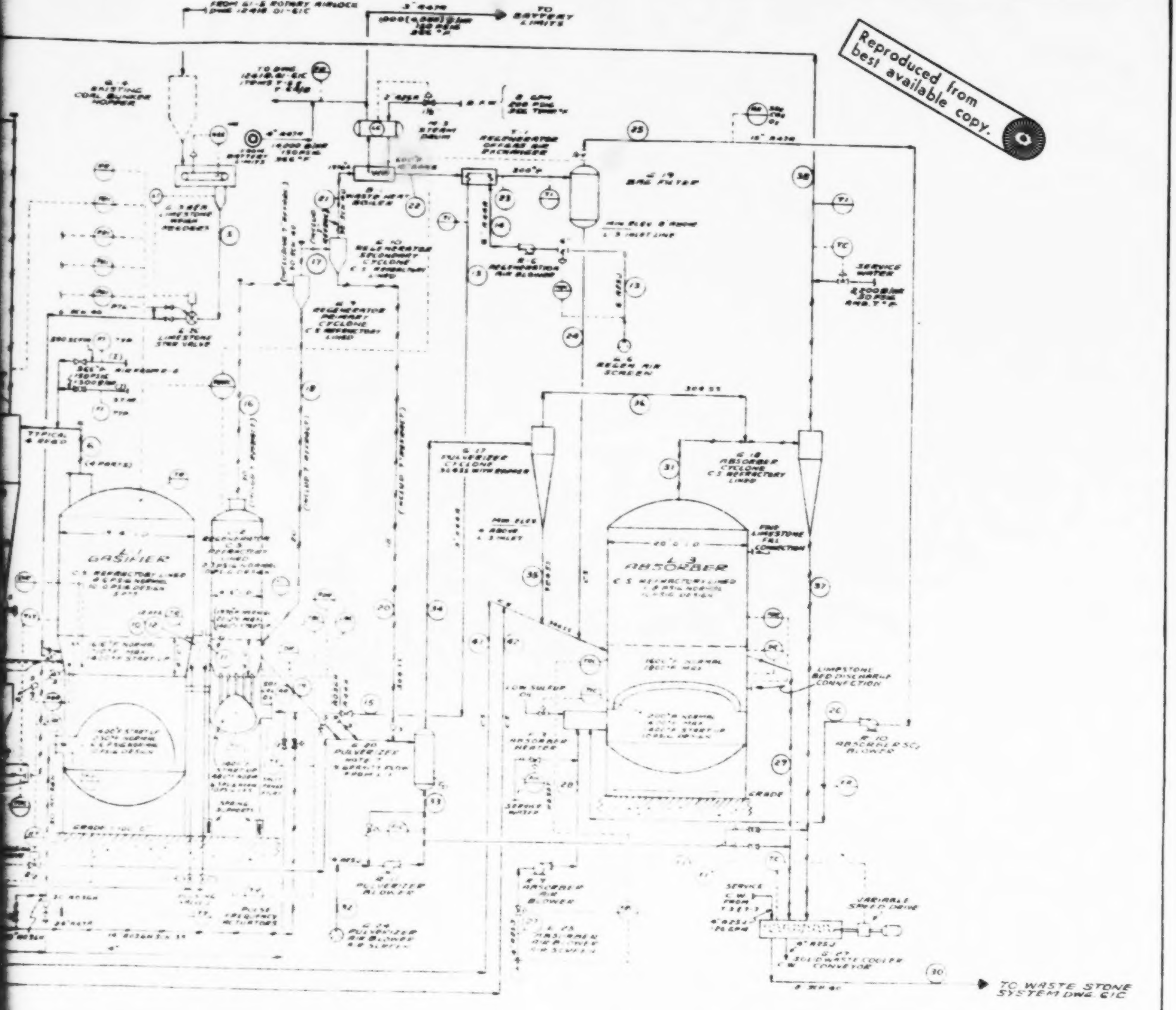
5 AMB = AMBIENT

6 N.C. = NORMALLY CLOSED

7 LINE FOR STREAM 19 TO BE 1 I.D. INCOLOY. LOCATE G 20 SUCH THAT SLOPED SECTION OF LINE WILL BE 2' MIN TO 6' MAX LENGTH @ 30° TO VERTICAL. HORIZONTAL RUN TO BE WELDED OR CAST PIECE OF MINIMUM LENGTH CONSISTENT WITH BRANCH CONNECTIONS FROM L-1 & 1C AND L-7. G 20 LOCATION TO ACCOMODATE STRAIGHT RUN CONNECTION TO L-1 @ 30° TO VERTICAL

8 AIR FLOW CONTROL TO B-8 TO BE RESET FOR CONSTANT RATIO TO HIGH SULFUR OIL FLOW (CONSTANT RATIO HAND ADJUSTABLE BY OPERATOR) WHEN FUEL OIL FLOW ABOVE 16,000 LB/HR. AIR FLOW TO B-8 TO BE RESET FOR CONSTANT COMBINED FLOW. B-8 DISCHARGE GAS TO L-1 WHEN FUEL OIL FLOW BELOW 16000 LB/HR

Reproduced from best available copy.


FROM 61-5 ROTARY AIRLOCK DWG 124181 01-61C
TO BATTERY LIMITS
EXISTING COAL BUNKER HOPPER
LIMESTONE WEIGH FEEDERS
LIMESTONE STAR VALVE
FROM BATTERY LIMITS
STEAM DRUM
WASTE HEAT BOILER
REGENERATOR OFFGAS AIR PREHEATER
REGENERATOR SECONDARY CYCLONE C.S. REFRACTORY LINED
REGENERATOR PRIMARY CYCLONE C.S. REFRACTORY LINED
REGENERATION AIR BLOWER
BAG FILTER
REGEN AIR SCREEN
SERVICE WATER
TYPICAL
GASIFIER
C.S. REFRACTORY LINED
REGENERATOR
C.S. REFRACTORY LINED
PULVERIZER CYCLONE
ABSORBER CYCLONE C.S. REFRACTORY LINED
FILL LIMESTONE FILL CONNECTION
ABSORBER
C.S. REFRACTORY LINED
LIMESTONE BED DISCHARGE CONNECTION
LOW SULFUR OIL
ABSORBER HEATER
PULVERIZER
ABSORBER SO2 BLOWER
GRADE
PULVERIZER BLOWER
ABSORBER AIR BLOWER
PULVERIZER AIR BLOWER AIR SCREEN
ABSORBER AIR BLOWER AIR SCREEN
PULSE FREQUENCY ACTUATORS
SPRING SUPPORTS
VARIABLE SPEED DRIVE
SOLID WASTE COOLER CONVEYOR
TO WASTE STONE SYSTEM DWG. 61C

PROCESS FLOW DIAGRAM

STONE & WEBSTER ENGINEERING CORPORATION

WESTINGHOUSE ELECTRIC CORP
FLUID BED OIL GASIFICATION
DEMONSTRATION PLANT
PROVIDENCE, RHODE ISLAND

REACTION SYSTEM
OIL GASIFICATION-
DESULFURIZATION CAFB

ISSUE	DATE	DESCRIPTION OF ISSUE	BY	CHK	APP
3	[illegible]	DELETE SCH FROM DUCTING	J.P.M.		
2	[illegible]	GENERAL REVISIONS BY WESTINGHOUSE	H.D.R.		
1	[illegible]	REFLECTS WESTINGHOUSE APPROVAL COMMENTS	H.D.R.		

J.O. NUMBER	DRAWING NUMBER	ISSUE
1241801	61A	3

Figure 7

and a regenerated stone purge stream at 297 kg/hr (655 lb/hr) are removed from the reaction system for processing in the stone processing system (dry sulfation process).

The gasifier and regenerator vessels are conventional refractory-lined fluidized bed vessels with carbon steel shells (shown in Figure 8). The gasifier has an inner diameter of about 5.95 m (19.5 ft). The design basis bed depth is 1.22 m (4 ft), and the freeboard height is about 3.7 m (12 ft). The air distributor is of a refractory construction to allow a start-up temperature below the grid of 760°C (1400°F). Fuel oil is injected into the gasifier bed at a level of about 152 mm (6 in) above the grid through three independent oil rings surrounding the vessel, which distribute oil to about 60 horizontal injection tubes, with one injection point per 0.47 m^2 (5 ft^2) of bed cross section. A single vertical baffle wall is placed in the bed and the regenerated stone inlet to the gasifier is situated on the side of the baffle opposite to the gasifier stone outlet to guarantee that the general circulation pattern will not bypass the bulk of the bed. The transport leg outlet is set at a 0.92 m (3 ft) bed depth and the inlet from the regenerator is set at a 152 mm (6 in) bed depth. Four fuel-gas outlets are arranged in the top of the gasifier.

The regenerator design is similar to that of the gasifier. The regenerator vessel has a 1.37 m (4.5 ft) diameter and is designed on the basis of a 1090°C (2000°F) operation temperature. No flow baffles are utilized in this regenerator. In addition to the circulating stone inlet and outlet, the regenerator has a waste stone drain line located in the wall at about 0.92 m (3 ft) bed depth.

Control

The major control points in the reaction section of the demonstration plant are the gasifier temperature and bed depth, and the regenerator temperature and bed depth. The gasifier temperature is controlled at about 870°C (1600°F) by the rate at which stack gas is recycled to the unit. The demonstration plant can also utilize steam or water injection as a method for gasifier temperature control. The gasifier bed depth is controlled by the rate at which stone is withdrawn from the regenerator through the waste stone line. The pressure differential across the gasifier bed controls the transport gas pulsing rate to this withdrawal leg.

The regenerator temperature is controlled by the rate of stone circulation between the gasifier and regenerator. The transport gas pulsing rate to the gasifier-to-regenerator transport leg is controlled by the regenerator temperature measurement. The regenerator bed depth is controlled by the regenerator-to-gasifier transport leg pulsing rate. An emergency quench system also exists for the regenerator in case of a temperature overshoot.

The performance of the gasifier and regenerator are not too sensitive to these control points so the control system response may be slow over a large operating range. Other important performance variables such as the gasifier sulfur removal efficiency and the regenerator product-gas sulfur dioxide and carbon dioxide contents will be monitored and controlled manually by setting the fresh limestone feed rate, the air-to-fuel ratio, the regenerator air rate, and the set points for the automatically controlled parameters.

Stone Processing System

Dry sulfation has been selected as a means of processing the waste stone from the reaction system in the demonstration plant. The dry sulfation stone processing system consists of the following components:

- An absorber
- A stone pulverizer
- Gas compressor
- Particulate removal components
- Heat exchangers.

The absorber, a fluidized bed reactor, is shown in Figure 9.

NOZZLES & MANHOLES

L-1 GASIFIER

SERVICE	MARK	SIZE	RATING	NOTES REMARKS
GAS TO CYCL.	1A-D	48"	150#	(4) REQ'D
FINES FROM CYCL.	4A-D	20"	"	(4) REQ'D SEE DETAIL
* OIL INJECTION	5	1/2"	"	(36) REQ'D
AIR & FLUE IN	6	30"	"	
SAMPLE CONN.	7A-F	16"	"	(6) REQ'D
DRAIN	8	18"		SEE DETAIL
VENT	9	16"	"	
PURGE GAS	11	1"	"	(36) REQ'D
MANHOLE	25A-C	36"	"	(3) REQ'D
THERMOWELL	30A-F	12"	"	(6) REQ'D
PRESS. CONN.	32A-C	12"	"	(3) REQ'D

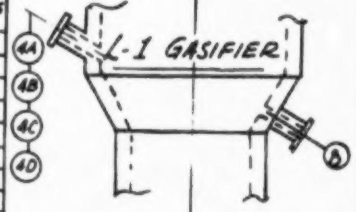
* SEE DETAIL

ESTIMATED WEIGHT - C.S.			78.000 LB.
"	"	S.S.	1.300 LB.
"	"	INCOLOY 800	3.700 LB.
"	"	REFRACTORY	212.000 LB.
"	"	ARCH	102.000 LB.
"	"	TOTAL *	397.000 LB.

L-2 REGENERATOR

SERVICE	MARK	SIZE	RATING	NOTES REMARKS
OFF GAS	1	30"	150#	(2) REQ'D
FINES	4	20"	"	
NET STONE OUT	5	20"	"	
AIR IN	7	24"	"	
VENT	9	20"	"	
SAMPLE CONN.	10	12"	"	
MANHOLE	25A & B	36"	150#	(2) REQ'D
THERMOWELL	30A & B	12"	"	(2) REQ'D
PRESS. CONN.	32A-C	12"	"	(3) REQ'D

ESTIMATED WEIGHT - C.S.			17.000 LB.
"	"	S.S.	500 LB.
"	"	REFRACTORY	45.000 LB.
"	"	ARCH	5.500 LB.
		TOTAL *	68.000 LB.

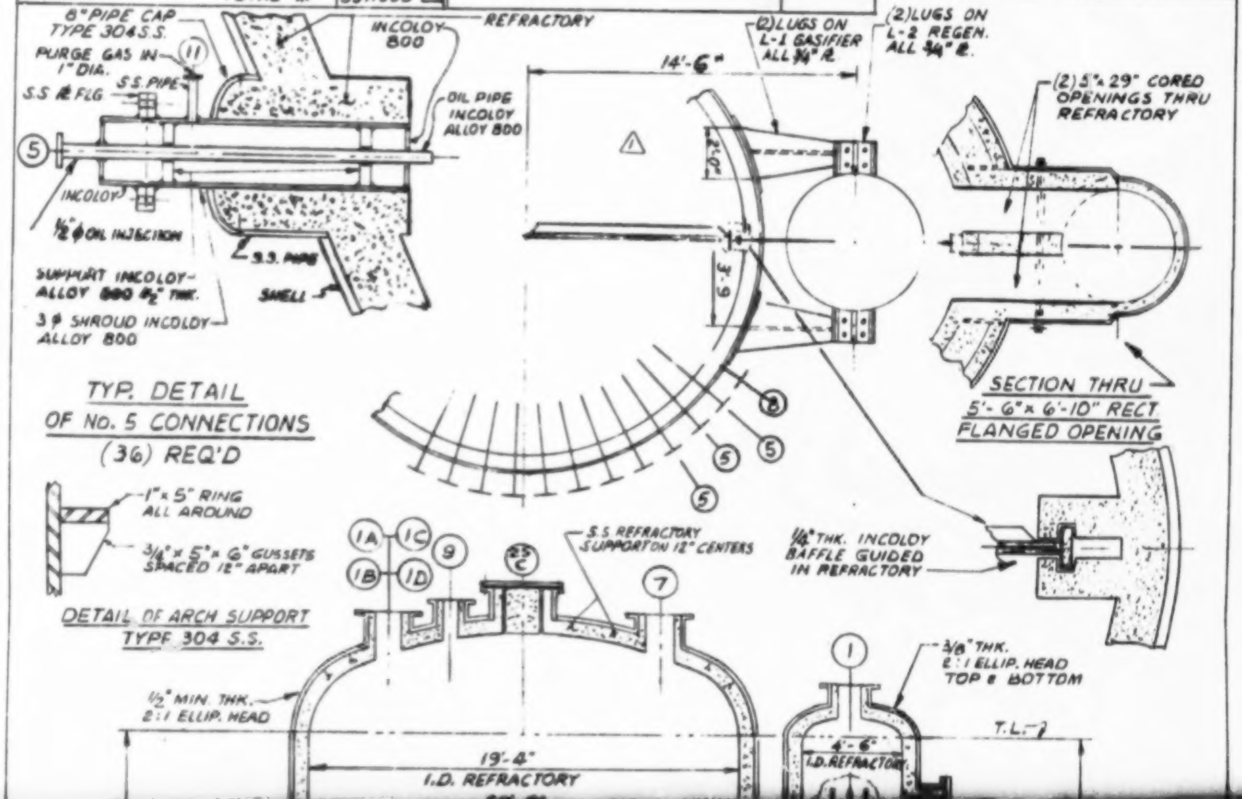

4A
4B
4C
4D
L-1 GASIFIER
B

LOCATION OF CONNS
(4A) (4B) (4C) (4D) & (B)

L-1 GASIFIER

NOTE FOR CONN. (4):
LOCATE (2) CONNECTIONS ON EITHER SIDE OF BAFFLE AS CLOSE TO THE REGENERATOR AS POSSIBLE.

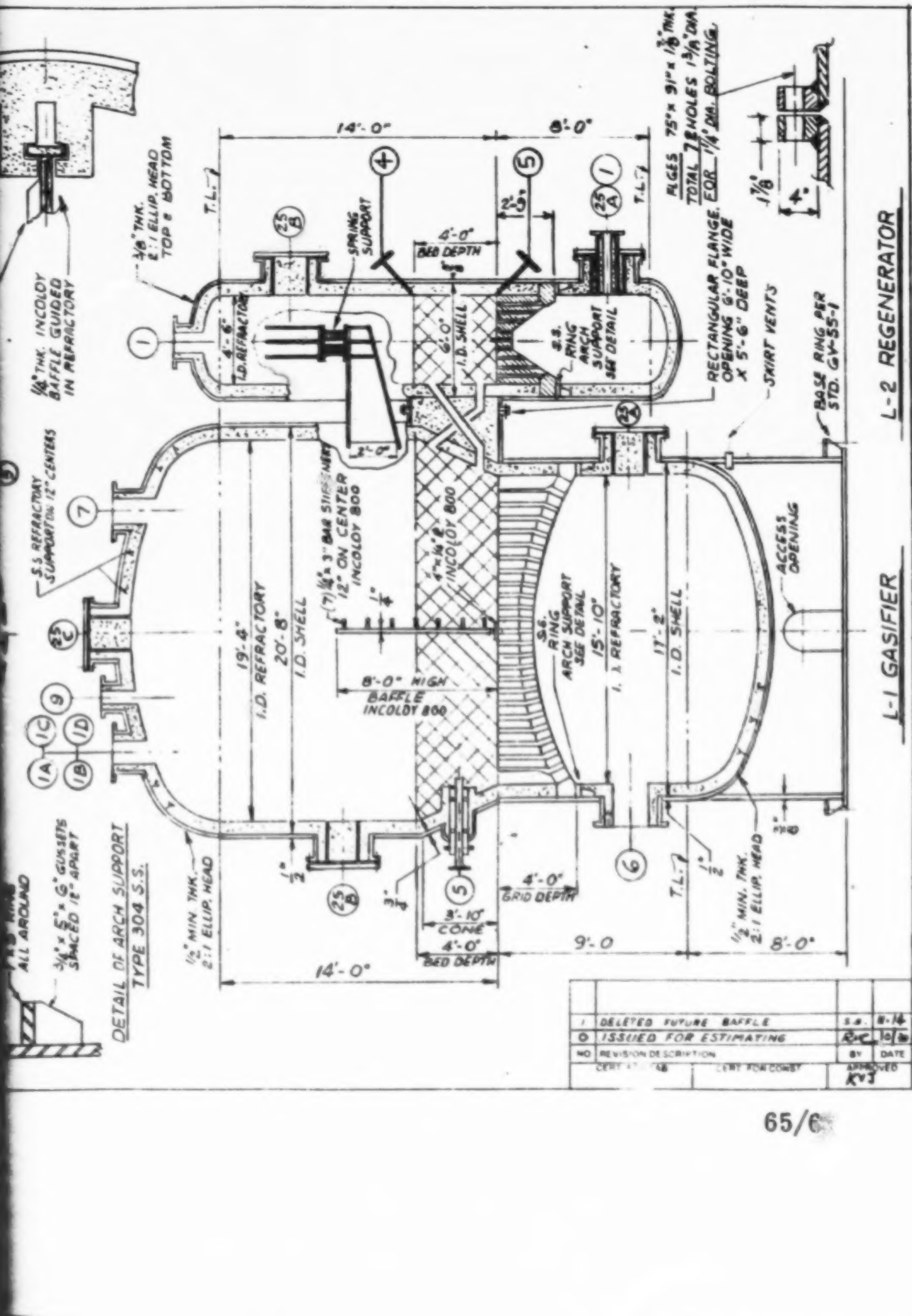
* BED NOT INCLUDED


8" PIPE CAP TYPE 304 S.S.
PURGE GAS IN 1" DIA.
11
S.S. PIPE
S.S. ℞ FLG.
5
INCOLOY
1/2" ϕ OIL INJECTION
SUPPORT INCOLOY ALLOY 800 1/2" THK.
3 ϕ SHROUD INCOLOY ALLOY 800
S.S. PIPE
SHELL
INCOLOY 800
REFRACTORY
OIL PIPE INCOLOY ALLOY 800
14'-6"
2'-0"
3'-9
(2) LUGS ON L-1 GASIFIER ALL 3/4" ℞
(2) LUGS ON L-2 REGEN. ALL 3/4" ℞.
(2) 5" x 29" CORED OPENINGS THRU REFRACTORY
B
5
5
5
1" x 5" RING ALL AROUND
3/4" x 5" x 6" GUSSETS SPACED 12" APART
1A
1B
1C
1D
9
25 C
7
S.S. REFRACTORY SUPPORT ON 12" CENTERS
1/4" THK. INCOLOY BAFFLE GUIDED IN REFRACTORY
1
3/8" THK. 2:1 ELLIP. HEAD TOP & BOTTOM
1/2" MIN. THK. 2:1 ELLIP. HEAD
T.L.
19'-4"
I.D. REFRACTORY
4'-6"
I.D. REFRACTORY

TYP. DETAIL OF No. 5 CONNECTIONS
(36) REQ'D

SECTION THRU 5'-6" x 6'-10" RECT. FLANGED OPENING

DETAIL OF ARCH SUPPORT TYPE 304 S.S.


DETAIL OF ARCH SUPPORT
TYPE 304 S.S.
ALL AROUND
3/4" x 5" x 6" GUSSETS SPACED 12" APART
1/4" THK. INCOLOY BAFFLE GUIDED IN REFRACTORY
S.S. REFRACTORY SUPPORT ON 12" CENTERS
3/8" THK. 2:1 ELLIP. HEAD TOP & BOTTOM
T.L.
14'-0"
8'-0"
4'-0" BED DEPTH
2'-3"
SPRING SUPPORT
4'-6" I.D. REFRACTORY
6'-0" I.D. SHELL
S.S. RING ARCH SUPPORT SEE DETAIL
2'-0"
FLGES 75" x 91" x 1 3/8" THK. TOTAL 72 HOLES 1 3/16" DIA. FOR 1 1/4" DIA. BOLTING
1 7/8"
4"
RECTANGULAR FLANGE OPENING 6'-10" WIDE X 5'-6" DEEP
SKIRT VENTS
BASE RING PER STD. GV-55-1
L-2 REGENERATOR
L-1 GASIFIER
(7) 1/2" x 3" BAR STIFFENER 12" ON CENTER INCOLOY 800
1/4" x 1/2" INCOLOY 800
19'-4" I.D. REFRACTORY
20'-8" I.D. SHELL
8'-0" HIGH BAFFLE INCOLOY 800
S.S. RING ARCH SUPPORT SEE DETAIL
15'-10" I.D. REFRACTORY
17'-2" I.D. SHELL
ACCESS OPENING
1/2" MIN. THK. 2:1 ELLIP. HEAD
4'-0" GRID DEPTH
3'-10" CONE
4'-0" BED DEPTH
9'-0
8'-0"
14'-0"
1/2"
3/4"
1
4
5
6
7
9
1A
1B
1C
1D
23
25A
25B
25C

GENERAL NOTES :

1. FOR MATERIAL AND DESIGN SPECIFICATIONS SEE DWG. Nº 12418-L-8A1

ISSUED FOR
ESTIMATING PURPOSE
ONLY

1	DELETED FUTURE BAFFLE	S.A.	8-14
0	ISSUED FOR ESTIMATING	Roc	10/30
NO	REVISION DESCRIPTION	BY	DATE
CERT. FOR BID	CERT. FOR CONST.	APPROVED KVJ	

CLIENT	WESTINGHOUSE			
LOCATION	PROVIDENCE RHODE I.			
SCALE —	STONE & WEBSTER ENGINEERING CORPORATION			
DESIGN	L-1 GASIFIER & L-2 REGENERATOR			
CHECKED				
DRAWN ZB	AREA	J.O. NO.	DRAWING NO.	ISSUE
CHECKED Roc		12418.01	12418 - L-1 & L-2-8A1	1

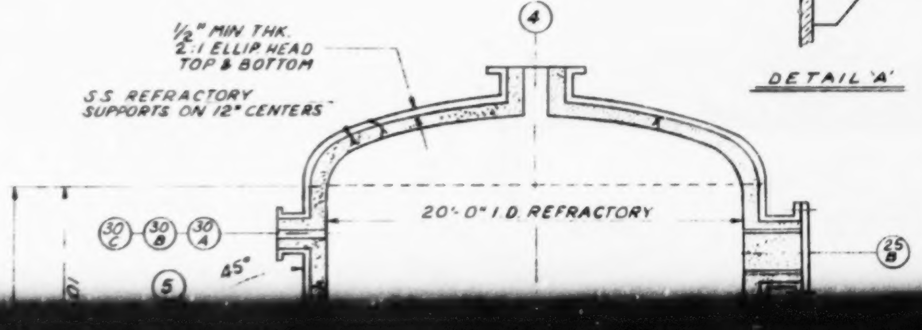
Figure 8

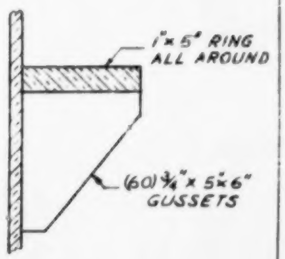
NOZZLES & MANHOLES

SERVICE	MARK	SIZE	RATING	NOTES REMARKS
GAS INLET	1	30"	150 #	
STONE OUTLET	2	18"	↓	
		-		
GAS OUTLET	4	24"	↓	
STONE FEED	5	18"	↓	
		-		
DUMP CONN.	7	24"	↓	COVER REFRACTORY LINED
MANHOLE	25A	36"	150 #	COVER REFRACTORY LINED
MANHOLE	25B	36"	↓	↓
THERMOWELL	30A	12"	150 #	
↓	30B	↓	↓	
↓	30C	↓	↓	
PRESS. CONN	32	16"	150 #	W/COVER SEE DWG

ESTIMATED WEIGHT - C.S. STEEL	78,000 LB
-"- -"- - S.S. STEEL	1,600 LB
-"- -"- - REFRACTORY	261,000 LB
-"- -"- ARCH	148,000 LB
-"- -"- TOTAL *	488,600 LB

* BED NOT INCLUDED


1/2" MIN THK.
2:1 ELLIP HEAD
TOP & BOTTOM
S.S. REFRACTORY
SUPPORTS ON 12" CENTERS
4
20'-0" I.D. REFRACTORY
30 C
30 B
30 A
5
45°
25 B
10


1"x 5" RING
ALL AROUND
(60) 3/4" x 5"x 6"
GUSSETS

DETAIL 'A'

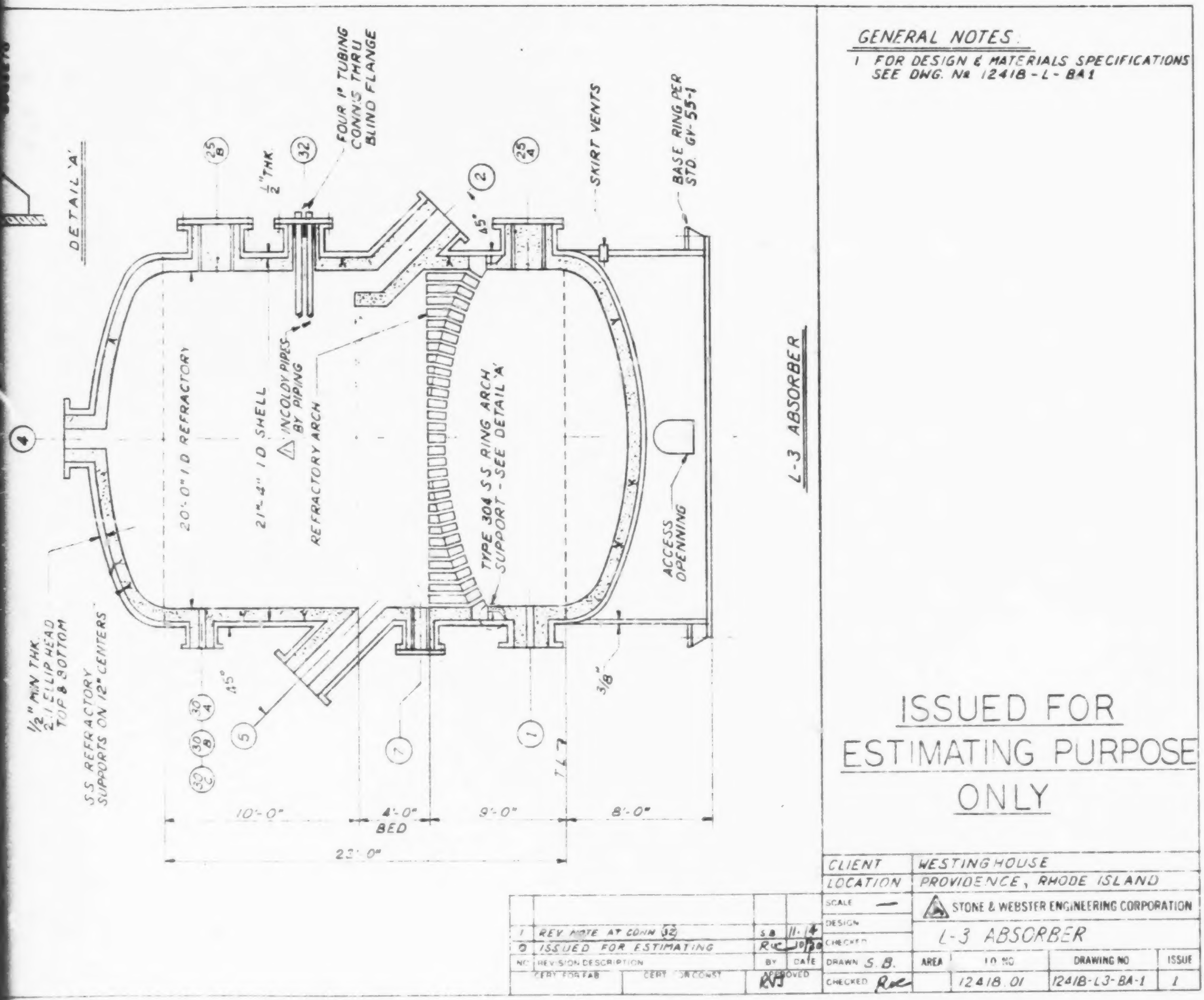

GENERAL NOTES:
1 FOR DESIGN & MATERIALS SPECIFICATIONS SEE DWG. № 12418-L-BA1
DETAIL 'A'
L-3 ABSORBER
FOUR 1" TUBING CONN'S THRU BLIND FLANGE
1/2" THK.
SKIRT VENTS
BASE RING PER STD. GV-55-1
45°
20'-0" I.D REFRACTORY
21'-4" I.D SHELL
INCOLOY PIPES BY PIPING
REFRACTORY ARCH
TYPE 304 S.S RING ARCH SUPPORT - SEE DETAIL 'A'
ACCESS OPENNING
1/2" MIN THK. 2:1 ELLIP HEAD TOP & BOTTOM
S.S. REFRACTORY SUPPORTS ON 12" CENTERS
3/8"
T.L.
10'-0"
4'-0"
BED
9'-0"
8'-0"
23'-0"
④
㉕ B
㉜
②
㉕ A
⑤
⑦
①
㉜ C
㉚ B
㉚ A
ISSUED FOR ESTIMATING PURPOSE ONLY
CLIENT
WESTINGHOUSE
LOCATION
PROVIDENCE, RHODE ISLAND
SCALE —
STONE & WEBSTER ENGINEERING CORPORATION
DESIGN
CHECKED
L-3 ABSORBER
DRAWN S.B.
CHECKED RC
AREA
J.O. NO
12418.01
DRAWING NO
12418-L3-BA-1
ISSUE
1
1 REV NOTE AT CONN ㉜ S.B. 11-14
0 ISSUED FOR ESTIMATING RC 10-30
NO REVISION DESCRIPTION BY DATE
CERT FOR FAB
CERT FOR CONST
APPROVED RVJ

Figure 9

The dry sulfation process contacts the waste stone from the reaction system and the boiler cyclone with the sulfur dioxide product-gas from the regenerator to produce a highly sulfated by-product stone and a low concentration sulfur dioxide-gas stream which is recycled to the gasifier. This contacting is carried out at about 870°C (1600°F) in a fluidized bed absorber. In order to enhance the rate of absorption and the degree of sulfation of the stone, the stone is pulverized while hot before being contacted with the sulfur dioxide. The dry sulfation process design specifications are based on the kinetic study presented in Appendix L. The sulfated by-product stone from the absorber is cooled and transported to the waste stone handling system.

Support Systems

The limestone, fuel-oil, and waste stone handling systems function to support the two major processing systems in the demonstration plant which have been described. These three support systems are conventional and, therefore, not described in any detail. The limestone handling system receives, stores, and feeds limestone at a controlled rate of about 1135 kg/hr (2500 lb/hr) to the gasifier. The fuel oil handling system receives, stores, and feeds high-sulfur residual oil to the gasifier at 12,700 kg/hr (29,000 lb/hr). The waste/by-product stone handling system transports and stores waste/by-product stone at about 1360 kg/hr (3,000 lb/hr).

Plant Turndown

The demonstration plant is designed to operate stably at specified levels of performance down to a partial load of 25 percent of full plant capacity. The stability of the reaction system is limited by the minimum operable fluidization velocity in the regenerator and gasifier. The minimum operable fluidization velocity of these units is one-half of the full capacity design velocity. Thus, turndown of the reaction system to 50 percent of full capacity can be carried out by a simple linear reduction by 50 percent in the capacity of all process streams. Further reduction in the plant output must be carried out by increasing the gasifier air-to-fuel ratio to such a degree that the 0.92 m/sec (3 ft/sec) minimum operating fluidization velocity is maintained.

The minimum operable regenerator fluidization velocity of 0.92 m/sec (3 ft/sec) is maintained by operating the regenerator with high excess air. This procedure may reduce the regenerator performance (product-gas sulfur dioxide concentration reduced and rate of calcium sulfate generation increased) but will not disrupt the plant performance significantly.

Turndown of the other components in the reaction system is facilitated by conventional practices.

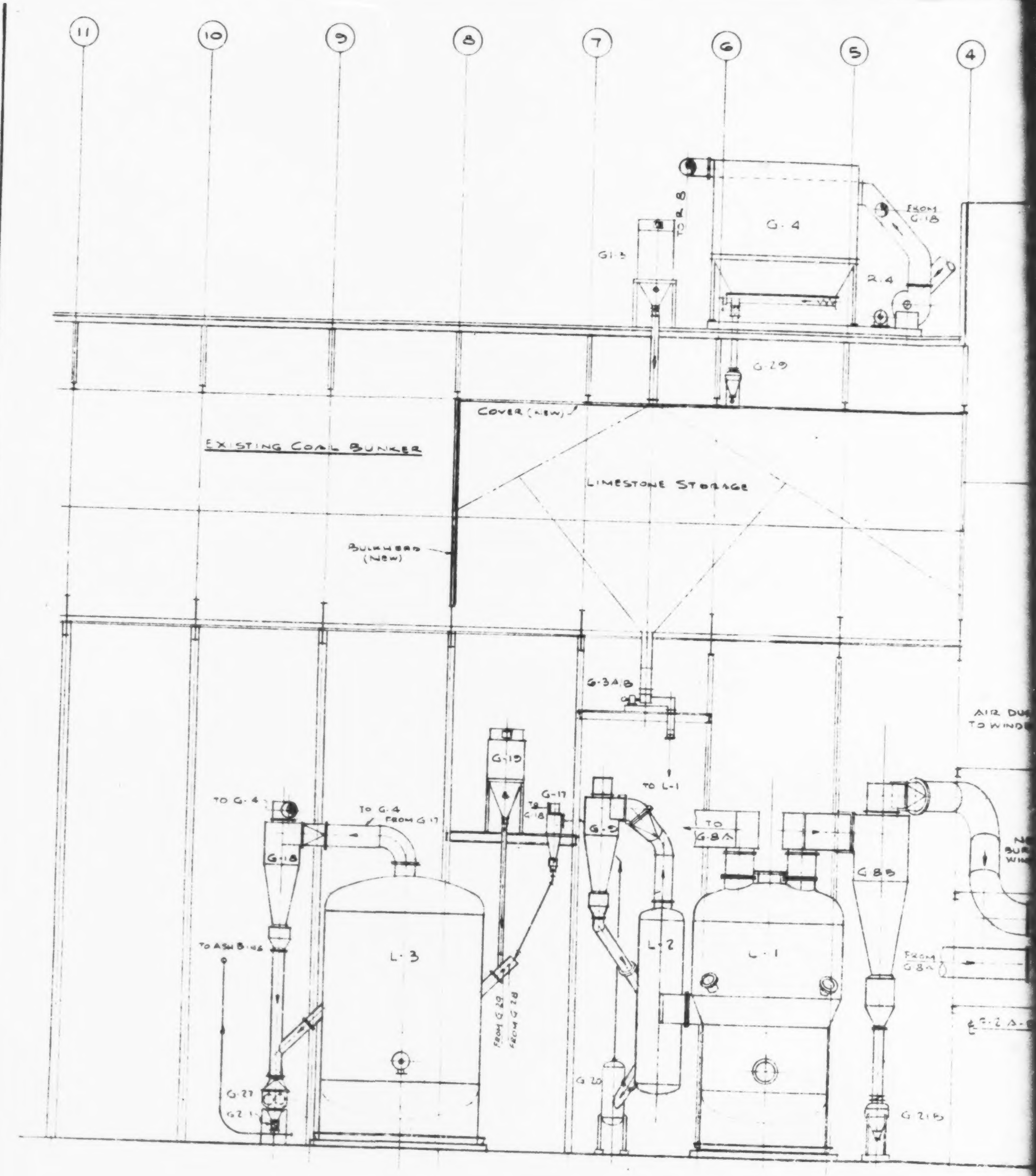
Control of Coke Deposition

The deposition of coke on the walls of the fuel-gas pipes and in the gasifier cyclones may be limited by proper design of the system and may be controlled by periodic burnout of the system. Smoothing all bends and transitions in the fuel-gas pipes reduces the formation of coke-lime deposits at these points. Based on pilot-plant performance (burnout required about every 200 hours of operation) the demonstration plant may require burnout about every 2000 hours of continuous operation. Demonstrated burnout procedures exist and are of the order of one percent of the time between burnouts. The probable plant operating schedule will be such that the gasification process may be burned out during scheduled plant shutdowns.

Detailed descriptions of the demonstration plant are given in Appendices B and E.

Plant Layout

Preliminary demonstration plant layout drawings are shown in Figures 10 through 13. The drawings, plan views and elevations, show all the major plant components, including the hot fuel-gas piping. Vessel L-1 is the gasifier; vessel L-2, the regenerator; vessel L-3, the absorber for


11
10
9
8
7
6
5
4
G-4
G1-3
TO R.B
FROM G-18
R-4
G-29
COVER (NEW)
EXISTING COAL BUNKER
LIMESTONE STORAGE
BULKHEAD (NEW)
G-3A.B
AIR DU
TO WINDE
G-19
G-17
TO G-18
TO L-1
TO G-4
TO G-4 FROM G-17
G-18
G-5
TO G-8A
G-8B
L-3
L-2
L-1
FROM G-8A
TO ASH BINS
FROM G-29
FROM G-28
G-27
G-21
G-20
G-21B

SECTION A-A

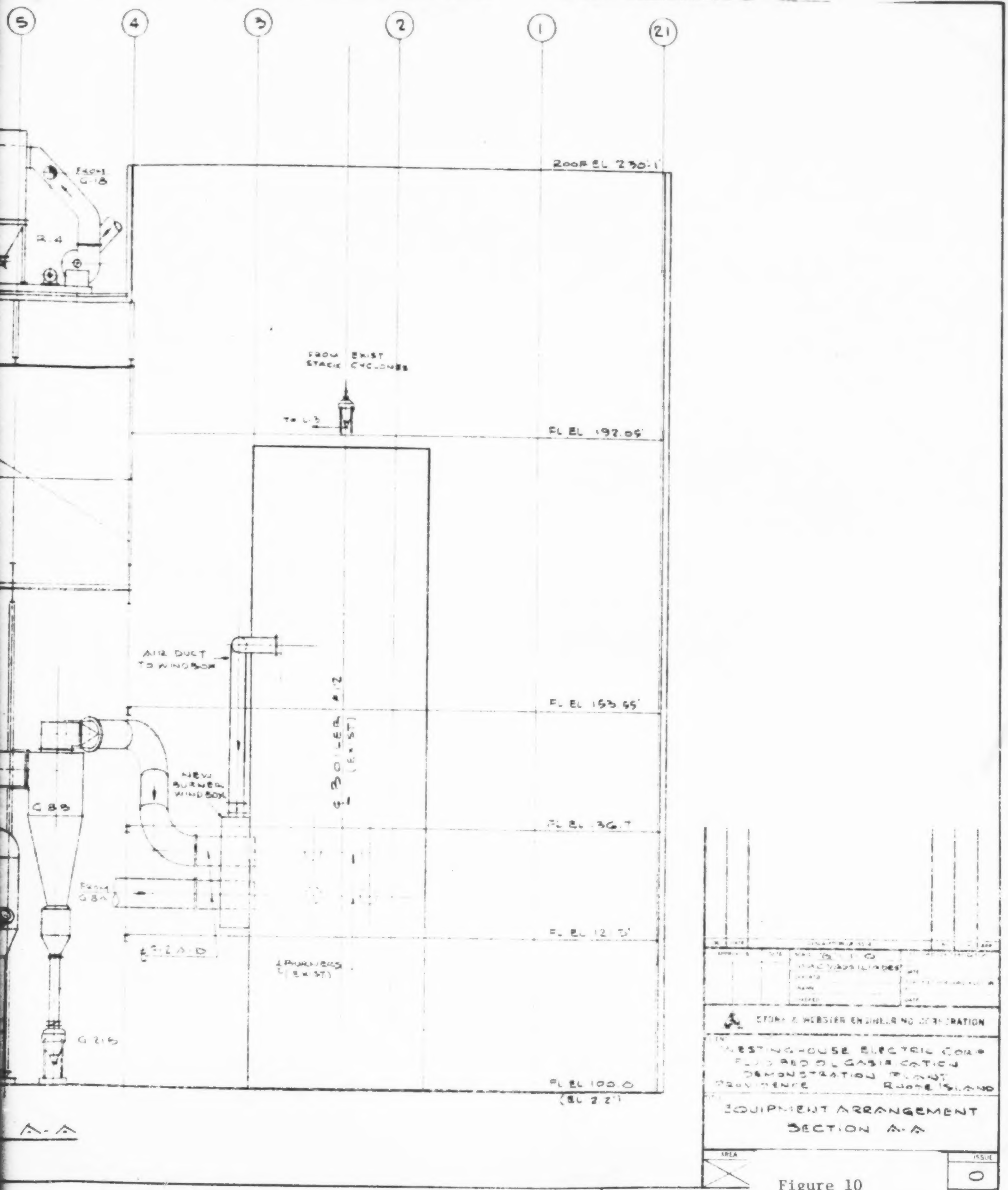
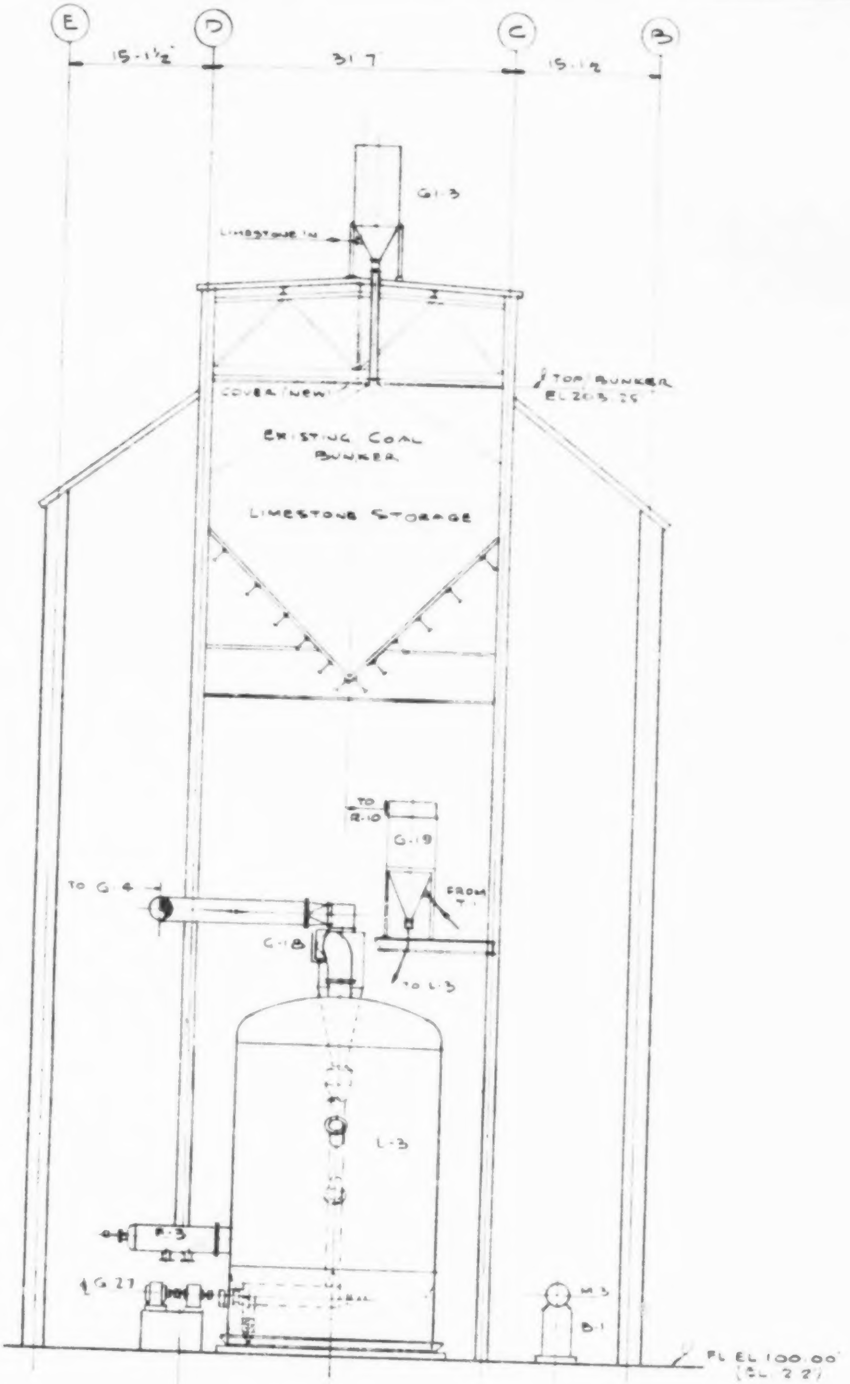
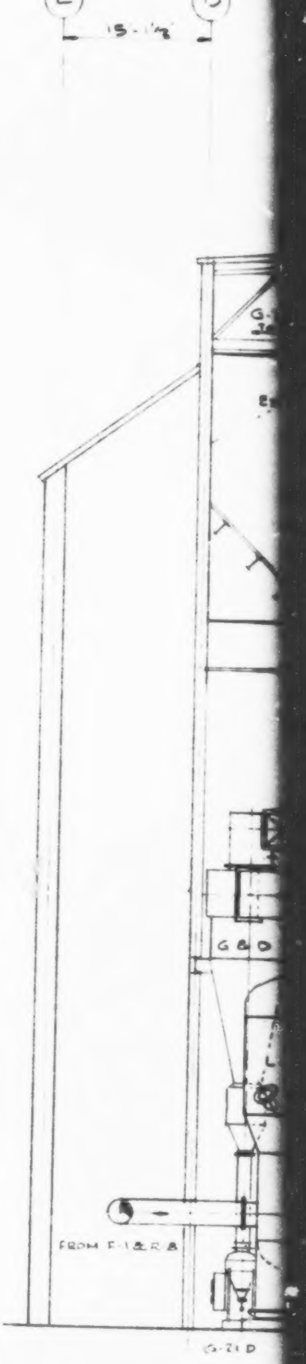

5
4
3
2
1
21
ROOF EL 230'-1"
FROM G-18
R-4
FROM EXIST STACK CYCLONES
TO L-3
FL EL 192.05'
AIR DUCT TO WINDBOX
BOILER #12 (EXIST)
FL EL 153.55'
NEW BURNER WINDBOX
C 8B
FL EL 136.7
FROM G 8A
FL EL 121.5'
E-2 A-D
BURNERS (EXIST)
G 21-B
FL EL 100.0
(EL 22')
A-A
STONE & WEBSTER ENGINEERING CORPORATION
WESTINGHOUSE ELECTRIC CORP
FLUIDIZED OIL GASIFICATION
DEMONSTRATION PLANT
PROVIDENCE RHODE ISLAND
EQUIPMENT ARRANGEMENT
SECTION A-A
AREA
ISSUE
0

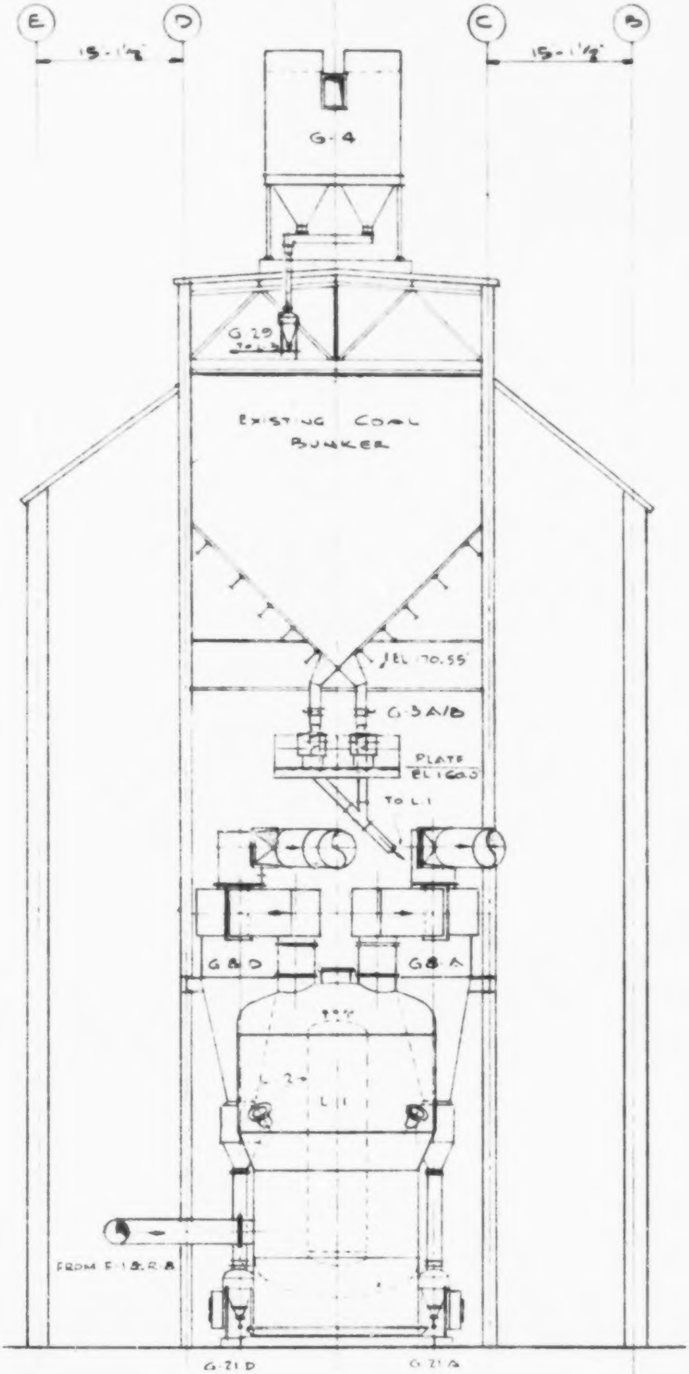
Figure 10


E
D
C
B
15-1½"
31-7
15-1½
G1-3
LIMESTONE IN
COVER (NEW)
TOP BUNKER
EL 203.25'
EXISTING COAL
BUNKER
LIMESTONE STORAGE
TO
R-10
G-19
FROM
T-1
TO G-4
G-18
TO L-3
L-3
P-3
G-27
M-3
B-1
FL EL 100.00'
(EL 22')

SECTION 'B-B'


E
D
15-1½"
G80
FROM F-1 & R-8
G-21 D

SE


E
D
C
B
15-1½"
15-1½"
G-4
G-29
TO L-1
EXISTING COAL BUNKER
EL 170.55'
G-3A/B
PLATE
EL 160.0
TO L-1
G8D
G8A
L-2
L-1
FROM F-1 B,R-B
G-21D
G-21A

SECTION C-C

NO.	DATE	DESCRIPTION OF ISSUE	BY	CHK	APP

APPROVALS	DATE		
		SCALE 8 : 1-0	CERTIFIED FOR FABRICATION BY
		DESIGN C VASSILIADES	DATE
		CHECKED	CERTIFIED FOR CONSTRUCTION BY
		DRAWN	
		CHECKED	DATE

STONE & WEBSTER ENGINEERING CORPORATION

CLIENT
WESTINGHOUSE ELECTRIC CORP
FLUID BED OIL GASIFICATION
DEMONSTRATION PLANT
PROVIDENCE, RHODE ISLAND

TITLE
EQUIPMENT ARRANGEMENT
SECTIONS B-B & C-C

AREA

Figure 11

ISSUE
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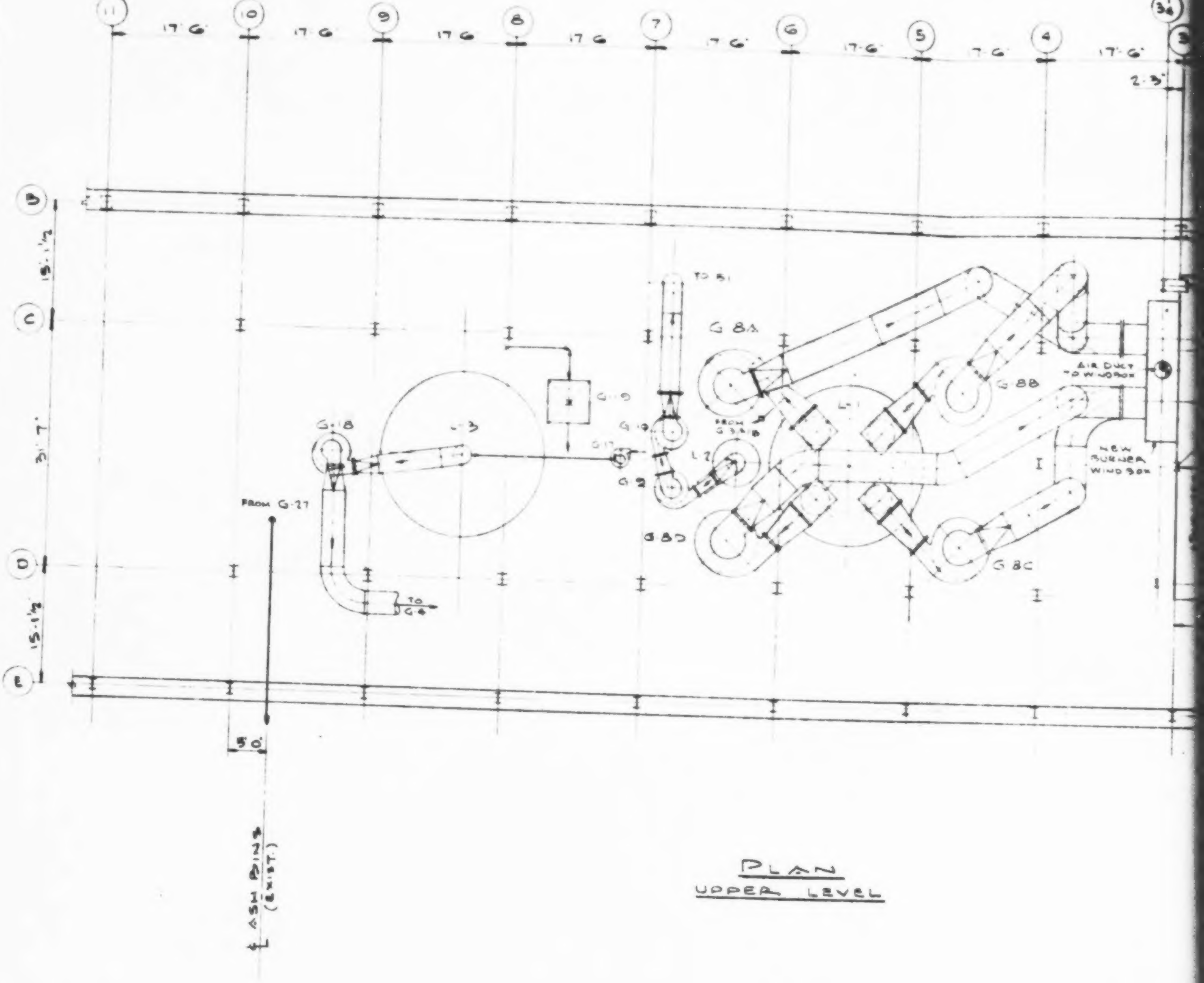
11
10
9
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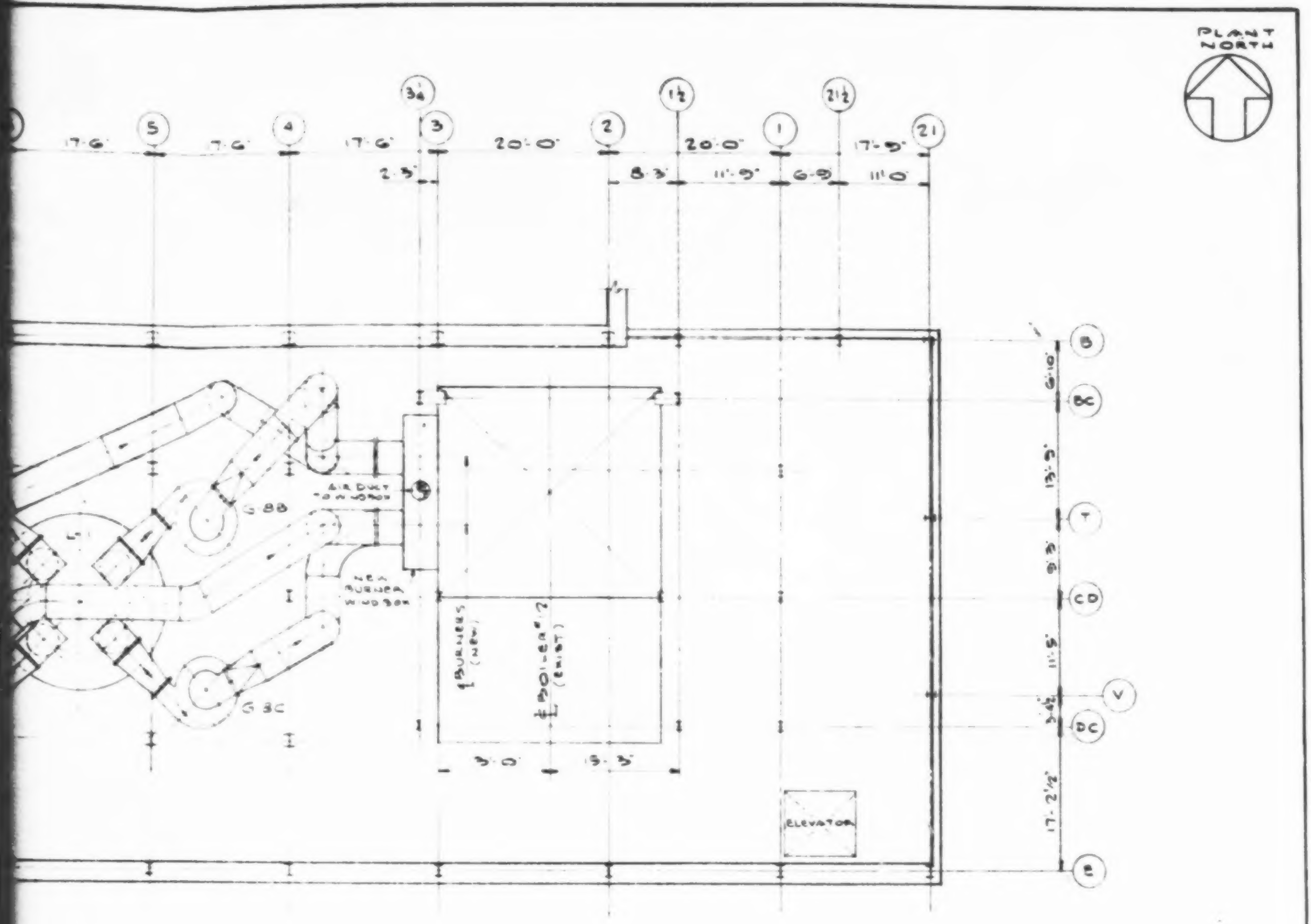
7
6
5
4
3½
3
17'-6"
17'-6"
17'-6"
17'-6"
17'-6"
17'-6"
17'-6"
17'-6"
2'-3"
B
C
D
E
15'-1½"
31'-7"
15'-1½"
TO B1
G 8A
G 8B
G 8C
G 8D
L-1
L-2
L-3
G-18
G-19
G-10
G-17
G-2
FROM G 3A/B
AIR DUCT TO WINDBOX
NEW BURNER WINDBOX
FROM G-27
TO G-4
5'-0"
ASH BINS (EXIST.)

PLAN
UPPER LEVEL


PLANT NORTH
5
4
3
3¾
2
1½
1
2½
21
17·6"
17·6"
17·6"
20·0"
20·0"
17·5"
2·3"
8·3"
11·9"
6·9"
11·0"
B
BC
T
CD
V
DC
E
6·10"
13·9"
9·3"
11·5"
3·4½"
17·2½"
AIR DUCT TO WINDBOX
G-88
G-8C
NEW BURNER WINDBOX
℄ BURNERS (NEW)
℄ BOILER #2 (EXIST)
3·0"
13·3"
ELEVATOR

AN
LEVEL

ISS	DATE	DESCRIPTION OF ISSUE	BY	CHK	APP

APPROVALS	DATE	SCALE ⅛" = 1'·0"	CERTIFIED FOR FABRICATION BY
		DESIGN C VASSILIADES	DATE
		CHECKED	CERTIFIED FOR CONSTRUCTION BY
		DRAWN	
		CHECKED	DATE

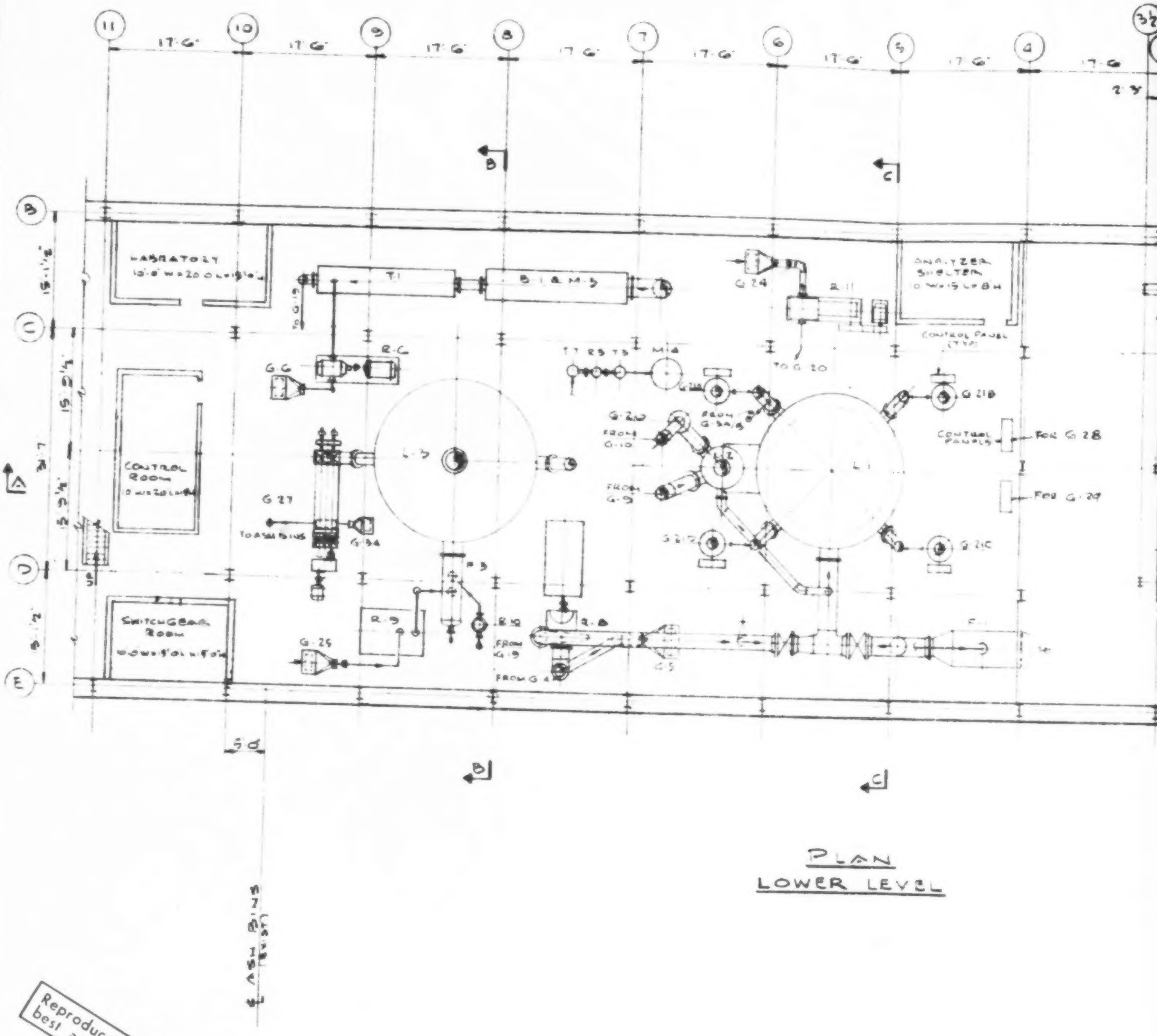
STONE & WEBSTER ENGINEERING CORPORATION

CLIENT
WESTINGHOUSE ELECTRIC CORP
FLUID BED OIL GASIFICATION
DEMONSTRATION PLANT
PROVIDENCE RHODE ISLAND

TITLE
EQUIPMENT ARRANGEMENT
PLAN
UPPER LEVEL

AREA
ISSUE 0

Figure 12


11
10
9
8
7
6
5
4
3½
17'-6"
17'-6"
17'-6"
17'-6"
17'-6"
17'-6"
17'-6"
17'-6"
2'-3"
B
C
B
C
A
B
C
C
J
E
15'-1⅛"
15'-2¼"
15'-7¼"
31'-7"
15'-1½"
5'-0"
LABRATORY
10'-0" W x 20'-0" L x 13'-0" H
CONTROL ROOM
10' W x 20' L x 14' H
SWITCHGEAR ROOM
10' W x 19' L x 15' H
UP
T1
B-1 & M-3
TO G-13
R-6
G-6
L-3
G 27
TO ASH BINS
G-34
R-3
R-9
R-10
FROM G-9
G-25
R-8
FROM G-4
G-5
T7 R3 T3
M-1A
G-21A
G-20
FROM G-10
FROM G-3A/B
L-2
FROM G-9
L-1
G-21D
G 24
R-11
TO G-20
ANALYZER SHELTER
10' W x 15' L x 8' H
CONTROL PANEL (TYP)
G-21B
CONTROL PANELS
FOR G-28
FOR G-29
G-21C
F-1
℄ ASH BINS

PLAN
LOWER LEVEL

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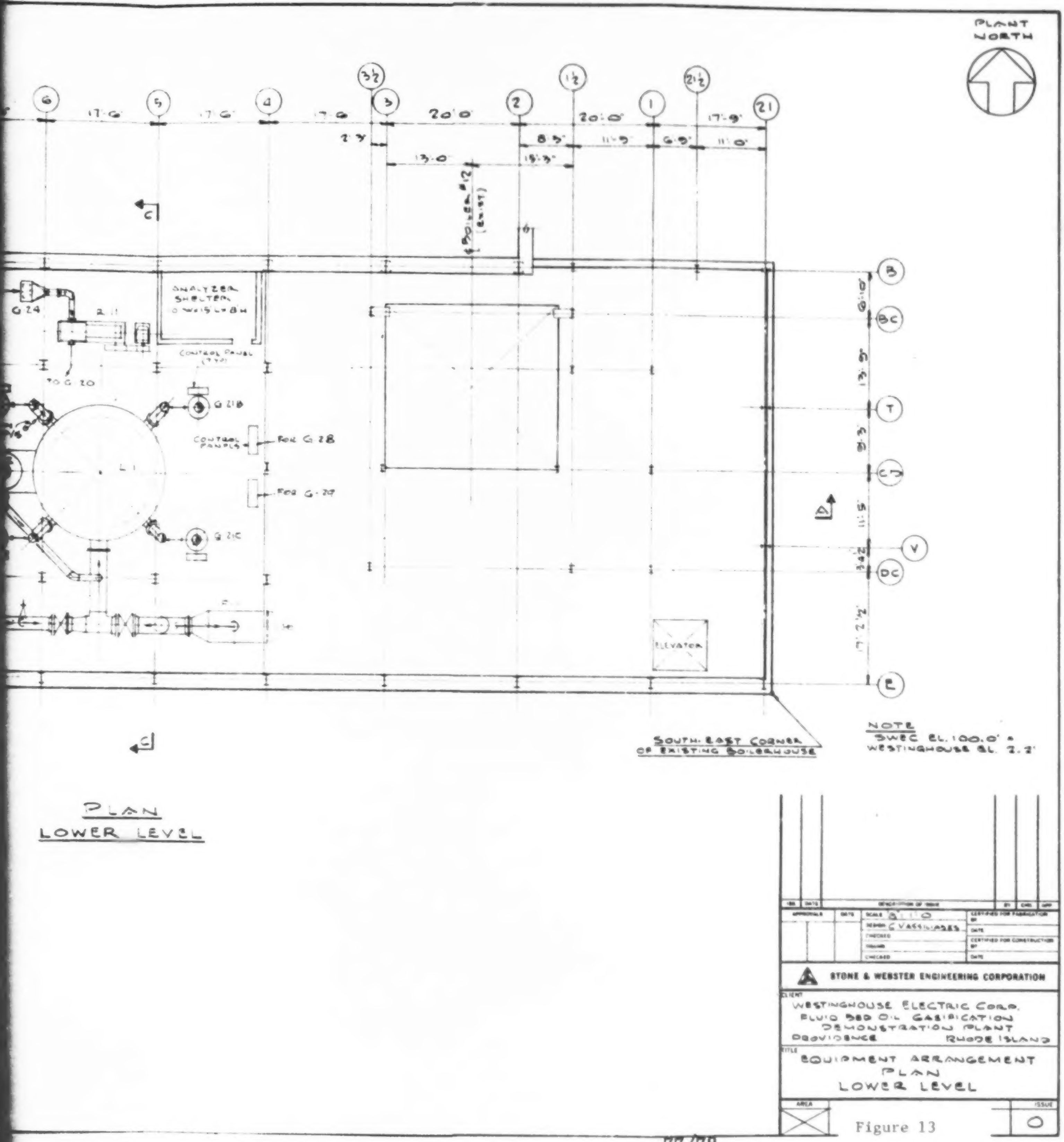
PLANT NORTH
6
5
4
3½
3
2
1½
1
2½
21
17'-6"
17'-6"
17'-6"
20'-0"
20'-0"
17'-9"
2'-7"
8'-5"
11'-5"
6'-9"
11'-0"
13'-0"
15'-3"
BOILER #12 (EXIST)
C
ANALYZER SHELTER
G-24
2-11
TO G-20
CONTROL PANEL
G-21B
CONTROL PANELS
FOR G-28
FOR G-27
G-21C
L-1
ELEVATOR
B
BC
T
C-D
V
DC
E
6'-10"
13'-9"
9'-5"
11'-5"
3'-4½"
17'-2½"
SOUTH-EAST CORNER OF EXISTING BOILERHOUSE
NOTE
SWEC EL. 100.0' = WESTINGHOUSE EL. 7.2'
PLAN
LOWER LEVEL
SCALE
C VASSILIADES
STONE & WEBSTER ENGINEERING CORPORATION
CLIENT
WESTINGHOUSE ELECTRIC CORP.
FLUID BED OIL GASIFICATION
DEMONSTRATION PLANT
PROVIDENCE RHODE ISLAND
TITLE
EQUIPMENT ARRANGEMENT
PLAN
LOWER LEVEL
AREA
ISSUE
0

Figure 13

stone processing; vessels G-8 (four of them), the gasifier cyclones; and vessels G-9 and G-19, the regenerator cyclones (in series). Vessel G-20 is the spent stone pulverizer and vessel G-17 is the pulverizer cyclone. The absorber cyclone is vessel G-18. The arrangement of the equipment has not been determined with regard to plant cost.

Boiler Modifications

An evaluation of the interfaces between the demonstration plant boiler and the fluidized bed residual oil gasification process was carried out. CE provided certain engineering services designed to establish parameters for firing the high-temperature, low heating-value gas in the CE steam generator (unit No. 12) at the Providence, Rhode Island plant of NEES (see Appendix C). Westinghouse provided CE with a design basis taken from conservative CAFB performance estimates (20 percent air/fuel ratio; high stack-gas recycle requirement).

An analysis of current performance data and latest equipment capabilities reveals that high-temperature, low heating-value gas can be fired in this unit. The maximum steam generating rate would be limited to 180,440 kg (400,000 lb) steam per hour with existing equipment because of I.D. fan limitations. To achieve a 204,120 kg (450.000 lb) per hour steaming rate new I.D. fans must be supplied. New burners will be required to fire the low heating-value gas. CE concluded that the following modifications should be made to the boiler (Appendix F presents further details of the evaluation):

- Tangential firing should be used. The base design using all four burners in one boiler face may lead to materials problems. Tangential firing provides advantages over single-face firing. Two of the four fuel-gas pipes would have to be extended about 15.2 m (50 ft) each for tangential firing capability.
- The unit side-walls must be modified for tangential firing.

- A feedwater line must be moved in order to insert air ducts, and so forth. Other plant/boiler components may require modification for access to the boiler.
- New air ducts for tangential firing must be added.
- A flame safeguard system must be provided for both the low heating-value gas and No. 6 oil (dual fuel boiler).
- New supports designed for 704°C (1300°F) gas temperature must be added at the lower back of the economizer. The normal gas temperature entering the economizer with oil-firing is about 593°C (1100°F). The economizer will be steaming with the low heating-value firing.
- In order to generate 204,000 kg/hr (450,000 lb/hr) of steam (the maximum boiler capacity) a new I.D. fan would be required (based on the gasification process base design conditions). Because of increased flue gas volume leaving the boiler (higher flue gas temperature) the existing I.D. fan would allow only about 182,000 kg steam/hr (400,000 lb steam/hr) to be generated (again based on the gasification process base design condition).
- Refractory plugs will be added to the existing horizontal burners to maintain present firing capability.

The study carried out by CE represents an initial evaluation of the boiler requirements. The study adopts the point of view that the plant should be modified to produce the maximum boiler design capacity, given certain assumptions about the gasification process. While this perspective is of interest, it may not be necessary or desirable to adopt this point of view for the demonstration plant. The study was also influenced by the basis which CE used for the gasification process.

The evaluation CE made was strongly influenced by a less than optimal understanding of the gasification process, its flexibility, and its unique conditions. Certainly, some modifications would be required to adapt the existing boiler to its role in a low heating-value gas demonstration plant. The economizer supports, burner plugs, and dual fuel safeguard system are examples of those required adaptations. In Westinghouse's opinion, however, the need for tangential firing, plus wall and tube modifications for that type of firing, is marginal. They are certainly the kind of modifications that would be required to maintain maximum boiler efficiency, but this is not a requirement of the demonstration plant. It appears that a substantial savings could be made by using the four-burners-in-one-wall approach to boiler firing, and with fewer potential problems than with longer, more convoluted fuel-gas pipes. If this approach caused some loss in boiler efficiency it would not be significant to the overall program. In addition, SWEC included in their costs new air ducts and wind boxes for the new gas burners. Two factors indicate that a new I.D. fan should not be required for the demonstration plant:

- The design for the plant is 195,000 kg steam/hr (430,000 lb steam/hr) (equivalent to 50 MW) and not 204,000 kg/hr (450,000 lb/hr).
- The base design conditions are the most conservative conditions and it is probable that lower flue gas volumes will exist in reality (by using a 17 percent air/fuel ratio, or water injection for gasifier temperature control, for example). Thus, the existing I.D. fan may be sufficient but, in any case, should not be replaced until the demonstration plant program has progressed to a point indicating its need.

Further review of the boiler performance and the recommended modifications is required before a decision can be made regarding material and equipment requirements.

DEMONSTRATION PLANT PERFORMANCE

The overall performance of the demonstration plant may be separated into three performance categories:

- Energy efficiency
- Environmental performance
- Technical performance of the components.

Energy Efficiency

The efficiency of the demonstration plant is dependent on three plant aspects:

- The fuel oil energy losses in the gasification process (thermal losses and carbon losses)
- The auxiliary power and steam required by the plant
- The efficiency with which the boiler utilizes the low heating-value fuel gas (stack losses).

The energy losses in the gasification process consist only of the energy required to heat and calcine fresh limestone feed and the negligible heat losses from the process equipment. The coke deposited on the gasifier bed is circulated to the regenerator, where it is combusted to raise steam as the regenerator sulfur dioxide gas passes through a waste heat boiler. The gas stream from the dry sulfation absorber is recycled to the gasifier. The losses from limestone calcination and heating amount to about one-half of a percent of the fuel oil energy input to the plant. Total losses from the gasification process would be expected to be about one percent.

The auxiliary power and steam requirements for the demonstration plant (based on the base design specifications) are summarized below (from Appendix E):

Steam - kg/hr (lb/hr): 1030 kPa (150 psig), 185°C (366°F)	
Produced	1810 (4,000)
Used (fuel oil feed system)	1410 (3,115)
Exported	401 (885)
Power - kw	2944

Steam is produced at a net rate of 401 kg/hr (885 lb/hr) for export from the hot regenerator-sulfur dioxide gas. The steam used by the plant is for the fuel oil system. Thus, there is no net steam consumption for the demonstration plant except for normal fuel oil system heating requirements. The auxiliary power requirement represents about 6 percent of the total power plant capacity for the base case conditions. Most of this power is associated with the compression of gasifier air and recycled stack gas. At an air/fuel ratio of 17 percent of stoichiometric the plant auxiliary power would be reduced to about 2,200 kw or about 4.5 percent of the power plant output. These power requirements are based on low efficiency blowers for gasifier air and stack-gas recycle (50 to 60 percent adiabatic efficiency). Since the boiler existing F.D. fan now handles only 80 percent of the normal combustion air quantity, some power savings would be realized over the normal F.D. fan power requirement.

CE (Appendix F) has estimated that (for the base conditions) the boiler efficiency with the low heating-value gas will be about 82.5 percent, as compared with about 87 percent with oil firing. This is due to higher stack-gas exit temperatures with the low heating-value gas. With a 17 percent of stoichiometric air/fuel ratio in the gasifier, the volume of gas passing through the boiler will be reduced, leading to higher boiler efficiency. The demonstration plant boiler will be dual-fired with No. 6 fuel oil to permit operation of the plant during periods of gasification process shutdown. While the normal steam superheat temperature of 510°C (950°F) can be achieved with the low heating-value gas, a steam temperature of only 482°C (900°F) can be achieved with No. 6 fuel oil. Maintenance of some damaged boiler baffles will increase the No. 6 fuel oil superheat temperature.

Environmental Performance

One of the major goals of the CAFB demonstration plant program is to demonstrate the ability of the process to utilize residual fuel oils and vacuum bottoms in an environmentally attractive way. The demonstration plant has been designed on the basis of the pilot-plant results summarized in Table 12. The environmental impact of the process is discussed in Appendix G.

New federal Source Performance standards for fossil fuel-fired steam generators of greater than 263.75 GJ/hr (250 million Btu/hr) heat inputs are 0.34 kg/GJ (0.8 lb/10^6 Btu) for sulfur dioxide and 0.43 kg/GJ (0.1 lb/10^6 Btu) for particulates. For the oil gasification demonstration plant, the sulfur dioxide output is projected to be 0.15 kg/GJ (0.35 lb/10^6 Btu) and is safely within the limit. This is based on 90 percent sulfur removal for the 2.6 percent sulfur residual oil for the demonstration plant. The total particulate emission is projected to be 0.0043 to 0.043 kg/GJ (0.01 to 0.1 lb/10 Btu), which meets the Performance Standard of 0.043 kg/GJ (0.1 lb/10 Btu).

Table 12

SUMMARY OF ENVIRONMENTAL PERFORMANCE OF
CHEMICALLY ACTIVE FLUIDIZED BED (CAFB) PILOT PLANT

Sulfur Removal Potential	-	Up to 95%
Sulfur Product	-	Sulfur fixed as solid calcium sulfate
NO_x Control Potential	-	Normal 270 ppm reduced to 155 ppm
Limestone Makeup Rate	-	0.5 to 1.5 times stoichiometric
Removal of Metals	-	100% fuel vanadium, 75% nickel, 40% sodium
Fuel Oils Utilized	-	Atmospheric residual (vacuum bottoms on batch unit tests)
Processing Steps Demonstrated	-	Gasification and regeneration

Since the particulate emission from the oil gasification plant consists mainly of calcium oxide, the environmental impact of calcium oxide particulate emission is also of interest. The stack exhaust from the oil gasification demonstration plant is equivalent to approximately 0.032 kg/453 kg (0.07 lb/1000 lb) effluent gas in the 50 MW demonstration plant, which is well below the emission standard of 0.1134 kg/453 kg (0.25 lb/1000 lb) for the cement industry.[16]

The rate of dry solids production from the demonstration plant dry sulfation system is about 1360 kg/hr (3,000 lb/hr). Several alternatives are being investigated for the disposal or utilization of this material (Appendix S): land disposal, ocean dumping, deep mine injection; or marketing for by-product recovery as gypsum, elemental sulfur, or sulfuric acid. Uses as construction material (concrete, aggregate, etc.) are also being explored.

The CAFB demonstration plant compares favorably environmentally with limestone and lime slurry scrubbers. Sulfur removal capabilities for the processes also appear comparable:

- CAFB provides a considerable reduction in nitrogen oxide emissions.
- The CAFB process consumes an order of magnitude less process water than slurry scrubbing processes.
- Limestone usage is comparable for the processes at about 1 mole of calcium per mole of sulfur removed from the fuel, but the CAFB process could potentially reduce this utilization to half that level by utilizing a regenerative operation with sulfur recovery.
- The CAFB process produces a dry product with potential market value rather than a sludge which is difficult to handle and requires large land areas for disposal ponds.

Demonstration Plant Process Performance

The technology which is being utilized in the demonstration plant preliminary design comes from three sources:

- Conventional chemical process plant practice
- Esso pilot-plant developments
- Laboratory bench-scale experimentation.

The process technical performance and reliability depends largely on the source of the technology. Many aspects of the demonstration plant design are based on improved plant reliability.

The provisions for oil feed, air feed, flue gas recycle, limestone feed, and waste stone handling are all by standard equipment in commercial use, and no abnormal problems are expected. Overcapacities of 15 percent on oil, 25 percent on gas blowers and baghouse, and 50 percent on limestone and waste stone are considered adequate to provide for variations from process design requirements. The maximum oil feed rate of 115 percent normal provides for 14,500 kg/hr (32,000 lb/hr) fuel, or 620 GJ/hr (588 MM Btu/hr) higher heating value to the system, or at 2 percent gasification heat loss, 615 GJ/hr (583 MM Btu/hr) to the boiler. In addition to larger than normal flue gas flow, the requirement of diverting 20 percent of combustion air to the gasifier with 80 percent through the air preheater to the boiler results in an abnormally high temperature from preheater to stack. Modifications to the preheater are under consideration but are not included in the cost estimate of this study.

The reliability and full capability of the boiler is protected by supplemental firing with the normal low-sulfur fuel oil.

Gasification

The pilot-plant work provides a sound basis for anticipation of 90 percent or better removal of sulfur at a level of 2.6 percent in the fuel. The depth of the bed of fluidized lime in the demonstration unit is more than double that in the pilot plant. This should provide a safety factor adequate to compensate for the degree of grossness in air and fuel distribution caused by practical mechanical designs and to allow for a modest recycle of sulfur dioxide from the absorber.

A possible source of difficulty is the formation of solid deposits of coke and lime dust on surfaces exposed to the hot fuel gas. It is believed that the rate of increase of depth of deposit is constant with time for given operating conditions. The effect of larger scale is beneficial in this respect since a depth which could shut down the pilot plant by restriction of small passages would have little effect in the large passages of the 50 MW unit. Pilot-plant runs have demonstrated simple and short duration burnout procedures. The utilization of vacuum bottoms has not been demonstrated in the Esso pilot-plant program; only limited batch runs have been carried out.

Regeneration

The pilot-plant work provides a sound basis for anticipation of sulfur rejection at 8 percent sulfur dioxide in the off-gas and loading of sulfur on the stone sufficiently low to provide pickup in the gasifier.

Sulfur Dioxide Absorber

No pilot-plant data on the absorber reaction has been generated, although tests in large Esso batch units are scheduled. The basis for the preliminary design has come from laboratory studies which Westinghouse has carried out (Appendix L).

Solids Transport

The circulation of solids between the gasifier and regenerator by a dense-phase pulsed transport scheme developed by Esso has proved to be a reliable circulation method, well suited to the CAFB process. Full-scale cold modeling of the demonstration plant circulation system will be carried out.

The Sturtevant systems for fines transport have been proved in all-metal systems at temperatures up to 390°C (750°F), but there is no experience with slug flow in refractory-lined pipe. In case of inoperability, substitution of Incolloy pipe is a possible modification.

COST ESTIMATE

SWEC has completed a preliminary cost estimate for the demonstration plant. The plant cost is given on the basis of end-of-1974 dollars and includes the fuel oil system, limestone handling system, waste stone handling system, reaction system, stone processing system (dry sulfation), and boiler modifications. The total plant cost does not include fee, contingency, start-up cost, and interest during construction. The total plant cost of $7,350,000 resulting from the SWEC study represents the cost for a demonstration plant with high excess capacity factors for many items of equipment and no optimization where high-cost plant components (such as the fuel-gas piping system) are involved. A breakdown of the plant cost is presented, and the contribution of the various equipment items and systems is discussed. The costs of some processing options and design modifications are also presented and discussed.

Demonstration Plant Cost Breakdown

The following tables summarize the demonstration plant capital investment. Table 13 presents the total SWEC estimated plant cost by general equipment classification. The high piping cost indicated in the table (65 percent of total process equipment) is due to the fuel-gas piping system.

Table 14 provides a cost breakdown according to the major plant systems -- reaction, stone processing, limestone handling, waste stone handling, and fuel oil:

- The reaction system in the table includes boiler modifications (fuel-gas burners and wind box) and plant buildings.
- The fuel-gas piping system (including supports and refractory) has been shifted from the process materials category as it appears in Table 13 to a process equipment item since it does represent unusual process piping.
- The limestone handling system includes all handling equipment from receiving up to the point of injection of limestone into the gasifier.
- The stone processing system accepts regenerator waste stone and the cooled and cleaned regenerator sulfur-dioxide gas and generates the waste/by-product stone.

The reaction system process equipment cost is about 65 percent of the total plant equipment cost, and the stone processing system is about 17 percent of the total plant equipment cost. The reaction system and stone processing system contribute about 77 percent of the total plant direct cost.

Table 15 summarizes the costs of the equipment items in the reaction system. The code identification for each item refers to the SWEC

Table 13

COST SUMMARY
(50 MW Plant)[a]

Item	Material ($)	Labor ($)	Total ($)
PROCESS EQUIPMENT			
A. Towers			
B. Boilers, Deaerators	32,000	600	32,600
F. Process furnaces	151,200	5,100	156,300
G. General equipment	775,000	45,900	820,900
L. Reactors	504,000	27,800	531,800
N. Drums	500	200	700
Q. Storage tanks	355,500	35,500	391,000
P. Pumps (including drives)	3,100	300	3,400
R. Compressors (including drives)	429,500	30,100	459,600
S. Stacks	--	--	--
T. Heat exchangers	58,700	1,800	60,500
TOTAL PROCESS EQUIPMENT	2,309,500	147,300	2,456,800
PROCESS MATERIALS			
C. Piping	1,048,000	545,000	1,593,000
D. Structures	169,800	48,800	218,600
E. Electrical	158,800	221,400	380,200
H. Buildings	24,500	25,400	49,900
J. Civil	65,800	110,100	175,900
K. Instruments	129,200	54,600	183,800
M. Insulation[b]	54,000	--	54,000
R. Painting[b]	87,900	--	87,900
Refractory lining[b]	60,000	--	60,000
TOTAL PROCESS MATERIALS	1,798,000	1,005,300	2,803,300
TOTAL DIRECT COST (DM*+DL)	4,107,500	1,152,600	5,260,100
DISTRIBUTABLE ACCOUNTS			
Y1. Insurance (excluding all risk)			
Y2. Federal and states taxes			
A. Temporary construction facilities			
X. Field office (including insurance and taxes)[a]			850,000
Z. Construction tools and equipment			
U. Other distributable items			
TOTAL DISTRIBUTABLE			850,000
SUBTOTAL-COST OF WORK			6,110,100
U. INDIRECT ACCOUNTS			
Engineering			
Design			
Other headquarters office			1,250,000
Taxes-headquarters payroll			
Overhead allowance			
TOTAL INDIRECT (HEADQUARTERS OFFICE)			1,250,000
TOTAL, PRESENT DAY PRICES (BARE COST)			7,360,100
Fee, Escalation, and Contingency			Not included
Order of Accuracy			15%

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[a] From original SWEC table.

[b] Subcontract.

Table 14

PLANT COST BREAKDOWN BY PLANT SYSTEM

	Demonstration plant system					
	Reaction[a] system	Stone processing system	Limestone[b] handling	Waste stone handling	Fuel oil[c] system	Total
Process Equipment						
Material	1,830,655	504,760	88,000	120,000	276,900	2,820,315
Labor	211,600	22,120	19,900	42,300	1,200	297,120
Total Process Equipment	2,042,255	526,880	107,900	162,300	278,100	3,117,435
Process Materials						
Material	1,003,085		36,500	22,600	214,400	1,276,585
Labor	521,000		31,300	63,300	239,700	855,300
Total Process Materials	1,524,085		67,800	85,900	454,100	2,131,885
TOTAL DIRECT COST	4,093,220		175,700	248,200	732,200	5,249,320

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[a] Includes major boiler modifications, plant buildings, and fuel-gas piping system (as a process equipment item).

[b] Includes limestone weight feeders and limestone star valve.

[c] Demonstration plant is dual-fueled with residual oil and low-sulfur oil.

Table 15

CAFB DEMONSTRATION PLANT

REACTION SYSTEM PROCESS EQUIPMENT COST

Item	Material($)	Labor($)	Total($)	Excess capacity
Gasifier (L-1)	222,000	13,400	235,000	Based on 20% air/fuel ratio; plant may operate at 17% air/fuel ratio
Regenerator (L-2)	41,000	900	41,900	
Gasifier Cyclones (G-8A-D)	213,000	1,800	214,800	30% volumetric flow excess
Gasifier Fines Recycle System (G-21-A-D), (G-22 A-D), (G-23), (G-23 A-D), (G-23 E-H)	188,100	5,800	193,900	About 40-minute storage capacity in lock hoppers - much larger than required
Regenerator Cyclones and Trickle Valves (G-9), (G-9A), (G-10), (G-10A)	47,600	900	48,500	15% volumetric flow excess
Gasifier Air/Flue Gas Blower and Screen (P-8), (G-5)	208,750	20,300	229,050	20% excess power
Regenerator Air Blower and Screen (P-6), (G-6)	39,230	1,800	41,030	30% excess volume: 50% excess power
Fuel-Gas Piping System	[illegible]1,415	150,000	671,415	About 20% excess pipe diameter
Regenerator SO_2-Gas System- Waste Heat Boiler (B-1), Steam Drum (M-1), Heat Exchanger (T-1), Bag Filter, Wrapper, and Discharge Valve (G-19), (G-19A), (G-19B)	58,300	1,500	59,800	
Flue Gas Recycle System-Stack Cyclone Lock Hopper, Valves and Connecting Piping (G-28), (G-30 A,B), (G-32), Recycle Flue Gas Baghouse, Screw Conveyor and Airlock (G-4), (G-4 A,B), Baghouse Lock Hopper, Valves and Connecting Piping (G-29), (G-31 A-B), (G-33), Flue Gas Booster Fan (P-4)	86,300	7,000	[illegible]3,300	About 25% volumetric flow excess: 25% power excess
Purge & Transport Gas System- PTG Blower Suction Cooler (T-7), PTG Blower (P-3), PTG Cooler (T-3), PTG Water Separator (M-4), PTG Filter (G-7)	61,460	3,400	64,860	About 25% excess volumetric flow, 35% excess power
Boiler Modifications-Fuel-Gas Burners (F-2A-D) and Windbox (F-24-D)	82,000	3,400	85,400	
Start-up Air Heater (F-1)	56,500	1,100	57,600	
Portable Chunck Trap Vessel (G-36)	5,000	300	5,300	
TOTAL PROCESS EQUIPMENT	1,830,655	211,600	2,042,255	

identifications appearing in the process flow diagram (Figure 7). The fuel-gas piping system equipment cost is about 33 percent of the total reaction system equipment cost at $671,000. This piping cost includes the shop-fabricated carbon steel pipe sections (1.17 m [46 in] and 1.22 m [48 in] diameters at a total length of 74.7 m [245 ft]), elbows, tees, reductions, flanges, welds, testing, bolt-up cost, hangers and supports, a 15 percent design allowance, refractory, and additional structures required for pipe system support. The table also lists the degree of overcapacity SWEC has specified for each equipment item.

Table 16 provides a similar breakdown for the equipment in the stone processing system. The major cost item in the stone processing system is the absorber, about 48 percent of the total equipment cost. More detailed cost breakdowns are included in Appendix E.

Assessment of Plant Cost

The estimated overcapacity (beyond the normal level of safety factor) built into many of the items listed in Tables 15 and 16 adds approximately $500,000 to the total plant cost. The single item having the largest overcapacity cost effect is the fuel-gas piping. The original design basis of 45.7 m/s (150 ft/sec) gas velocity (Appendix D) has been reduced by overcapacity to less than 30.1 m/s (100 ft/sec), resulting in about a 21 percent increase in piping diameter. The layout of the fuel-gas piping system and the gasifier location has not been optimized for the preliminary design. Potentially, large cost reductions could be realized by optimization. But, since the SWEC design basis places the four burners in a single face of the boiler, a plant with tangential burners at the four corners of the boiler would result in a more complex and costly fuel-gas piping system (possibly as much as $300,000 additional equipment cost for 30.1 m [100 ft] additional length). The need for tangential burners has not been definitely established, although CE has recommended it.

The gasifier is not an extremely important cost item in the total plant. Operating the gasifier at 1.22 m/s (4 ft/sec) velocity

Table 16

CAFB DEMONSTRATION PLANT

STONE PROCESSING SYSTEM (DRY SULFATION) EQUIPMENT COST

Item	Material($)	Labor($)	Total($)	Excess capacity
Absorber (L-3)	241,000	13,500	254,500	
Absorber Cyclone and Trickle Valve (G-18), (G-18A)	24,000	300	24,300	35% excess volumetric flow
Jet Pulverizer (G-20)	34,200	560	34,760	45% excess volumetric flow
Pulverizer Cyclone and Trickle Valve (G-17), (G-17A)	7,600	300	7,900	27% excess volumetric flow
Absorber Air Blower and Screen (P-9)	33,130	1,200	34,330	15% excess flow: 65% excess power
Absorber SO_2 Blower (R-10)	67,000	1,700	68,700	15% excess flow: 15% excess power
Pulverizer Air Blower and Screen (R-11), (G-24)	37,730	1,800	39,530	5% excess flow: 37% excess power
Solid Waste Cooler/ Conveyor (G-27)	47,500	2,200	49,700	50% excess flow
Absorber Heater (F-3)	12,600	560	13,160	
TOTAL PROCESS EQUIPMENT	504,760	22,120	526.880	

rather than at the 1.83 m/s (6 ft/sec) design velocity which has been applied would result in a total plant cost increase of only about $120,000. Pilot-plant studies indicate that a 0.61 m (2 ft) bed depth may be sufficient for the achievement of 90 percent sulfur removal, resulting in a 0.61 m (2 ft) decrease in bed depth from the base design. On the basis of pilot-plant results, this decrease would substantially reduce the bed and gasifier distributor pressure drop and could reduce the fines elutriation rate (with a 1.22 m/s [4 ft/sec] velocity) by an order of magnitude. Elimination of the gasifier cyclones and fines recycle system, if possible, would result in a total plant cost reduction of about $900,000 and an additional reduction in fuel-gas piping system complexity.

The preliminary design of the demonstration plant does not represent a prototype of a commercial retrofit system. Optimization of the plant (piping system layout, gasifier location and operating conditions, and so on) and elimination of excess capacity would substantially reduce the demonstration plant cost. The piping system represents the largest cost and the most sensitive cost item in the plant.

Processing Options

In addition to the demonstration plant concept which was selected for the preliminary design study (regenerative process with dry sulfation of the waste/by-product), two other process concepts could be investigated in a demonstration plant -- once-through operation and regenerative operation with sulfur recovery. From the initial design study (Appendix I) and the results of the final preliminary study on the dry sulfation processing concept the cost of a demonstration plant using sorbent regeneration, recovery of sulfur from the regenerator sulfur dioxide, and slurry carbonation of the waste/by-product material is estimated to be about $10.5 million; the sulfur recovery process is estimated to be about $3 million; the reaction system and support systems are assumed to cost about the same as in the SWEC study; and the slurry carbonation system cost is about the same as that of the dry sulfation system. This cost would be reduced by the same factors discussed for the final preliminary design (in other words, optimization and overcapacity reduction).

Once-through operation could result in a low-cost demonstration plant if two assumptions were valid—very high lime utilization (more than 75 percent) can be achieved and the highly utilized calcium sulfide waste can be marketed without further processing. Assuming the validity of these propositions, the cost of the demonstration plant would be about $5.8 million without additional optimization or excess capacity reduction. With optimization of the plant (elimination of the gasifier cyclones and fines recycle systems; reduction in piping system length and complexity) and reduction in the excess capacity, the once-through demonstration plant could cost as little as $4.5 million.

The processing concept using regenerative operation with the sulfur recovery system used in the initial design study did not appear attractive from a cost aspect. Other processes being developed, however, may render this option economically competitive. Once-through processing is potentially attractive from a cost aspect, but the assumptions of high sorbent utilization (Appendix R) and direct disposal of calcium sulfide (Appendix S) require extensive investigation.

Operating Cost

SWEC has estimated the demonstration plant annual operating costs. Unit costs used are as provided by NEES, basis 1976; internal power 23 mills/kWh; No. 6 fuel oil; $2.25/MM Btu (HHV); capital charges, 20 percent of bare cost/year; maintenance (including construction tools, supplies and labor), 5 percent of bare cost/year.

Annual costs are based on 7000 hr/yr stream operation.

Bare Cost		$7,360,100.00
Direct Operating Costs		
Power	$474,950.00	
Operating labor	304,500.00	
Limestone supply	76,250.00	
Solid Waste disposal	40,880.00	

Maintenance	368,005.00	
Total		1,264,585.00
Indirect Operating Costs		
Capitol charges		1,472,020.00
Total Direct & Indirect Costs		2,736,605.00

This annual operating cost amounts to a fuel adder of about 76¢/MM Btu or 7.8 mills/kWh. For an optimized plant with overcapacity reduced, the fuel adder could be substantially reduced. Operation of the plant with a cheaper vacuum bottoms fuel, if available, could result in a fuel credit relative to the cost of No. 6 fuel oil.

VI. DETAILED, DESIGN, CONSTRUCTION, AND OPERATION

A scope of work, schedule, and cost estimate are presented here for the Phase II work, detailed design, and construction. Shake-down and commissioning operations and an experimental test program projected for the plant are reviewed. Implementation of the second phase of the program, however, detailed design and construction, is contingent on the decisions reached by the project participants after evaluation of the Phase I work.

SCOPE OF WORK

Selection of Architect-Engineer

A list of qualified architect-engineers will be prepared. Specifications and procurement documents based on the preliminary design will be prepared and distributed to the candidate architect-engineers. Proposals for the detailed design and construction, and cost estimate for operation (Phase III), will be received and evaluated and an architect-engineer selected.

Detailed Engineering, Procurement, and Construction

The services of the architect-engineer will include but not be limited to the following categories of work:

- Preparation of layouts, architectural drawings, and flow diagrams to reflect changes from the preliminary design caused by size and specifications of equipment finally selected
- Completion of mechanical equipment specifications and vessel drawings to show actual sizes and orientation of all interconnections

- Preparation of structural detail and foundation drawings for buildings and equipment
- Provision of a piping model plus piping orthographics for all unmodeled piping
- Provision of electric lighting, power wiring, and heat tracing installation drawings, including motor control center layouts
- Provision of instrument sizing and selection, instrument installation details, and final panel layout drawings
- Provision of ten copies of all manufacturers' parts lists, spare parts recommendations, operating instructions, and shop drawings
- Provision of critical path scheduling to monitor on a bimonthly basis the timely performance of the work
- Provision of a Division of Work and a Mechanical Completion Definition to allocate responsibility between Westinghouse, NEES, and the architect-engineer for all phases of the work
- Fabrication and/or procurement of all necessary materials, equipment, and fabrication services to complete the plant
- Performance and completion of all construction and installation work necessary to erect the demonstration plant at the Manchester Street Station, Narragansett Electric Company, Providence, R.I.
- Receiving and storing of all material and equipment, and procurement and furnishing of all manual labor, supervision, and subcontractor services necessary to perform the work to the best advantage.
- Arranging for the provision of all utilities and services for the construction and testing of the facility, payment of all applicable taxes and obtaining of all necessary permits.
- Connecting the demonstration plant into the existing electrical, steam, water, sanitary sewage, compressed air, natural gas, and storm sewer systems at battery limits

- Performance of pressure testing of piping and vessels, testing of mechanical functioning of the equipment, and final calibration of the instruments as a minimum, prior to certification of mechanical completion.

SCHEDULE

The overall program, as shown on the computerized critical path plot figure, is projected to require three years to complete. The value of the critical path scheduling technique is that it enables priorities to be established on the basis of material lead times and provides a logical sequence for interconnecting activities. When looking at Figures 14 and 15, one can see each activity described above a line between two boxes with numbers.

The lead box is called the predecessor and the follow box is called the successor event, as shown on the critical path listing attached.

The information below the line is the estimated duration of the specific activity in weeks. The most critical path is located on the top of the page, and subsequent paths are listed by their criticality.

The durations for the various activities shown on the schedule (Table 17) are assumptions generally valid for labor efficiencies and for labor and material availabilities prevailing in the fall of 1974. Westinghouse expects that individual activity durations may vary considerably from some of those shown on the schedule. This is, however, unlikely to affect significantly the overall time required to complete the demonstration plant construction.

COST ESTIMATE

An estimate tabulation is presented in Table 13 to show the detailed costs of the oil gasification system with dry stone sulfation installed at the Manchester Street Station, including new high-sulfur fuel oil tankage, piping and pumps, new hot-fuel gas burners, and new limestone

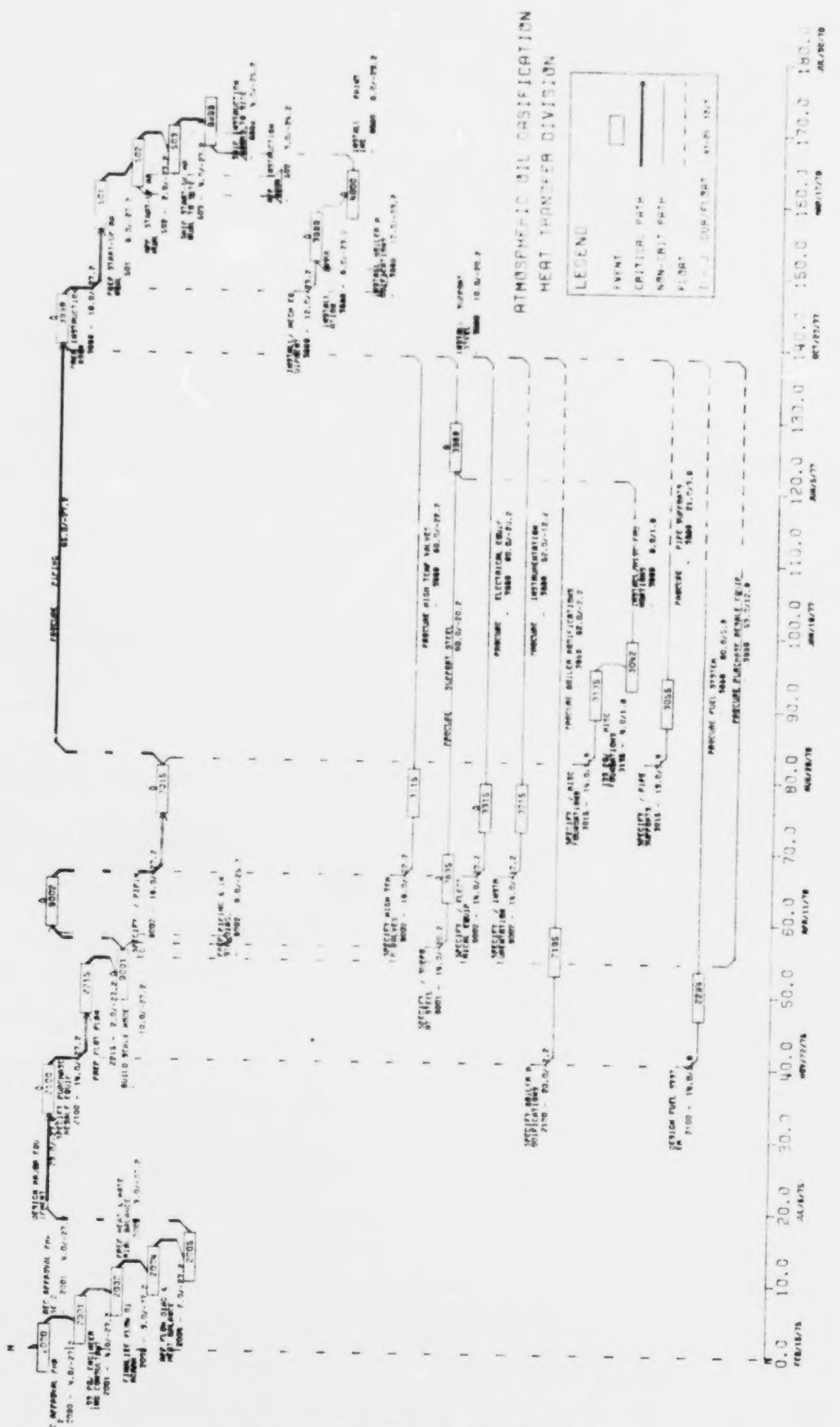
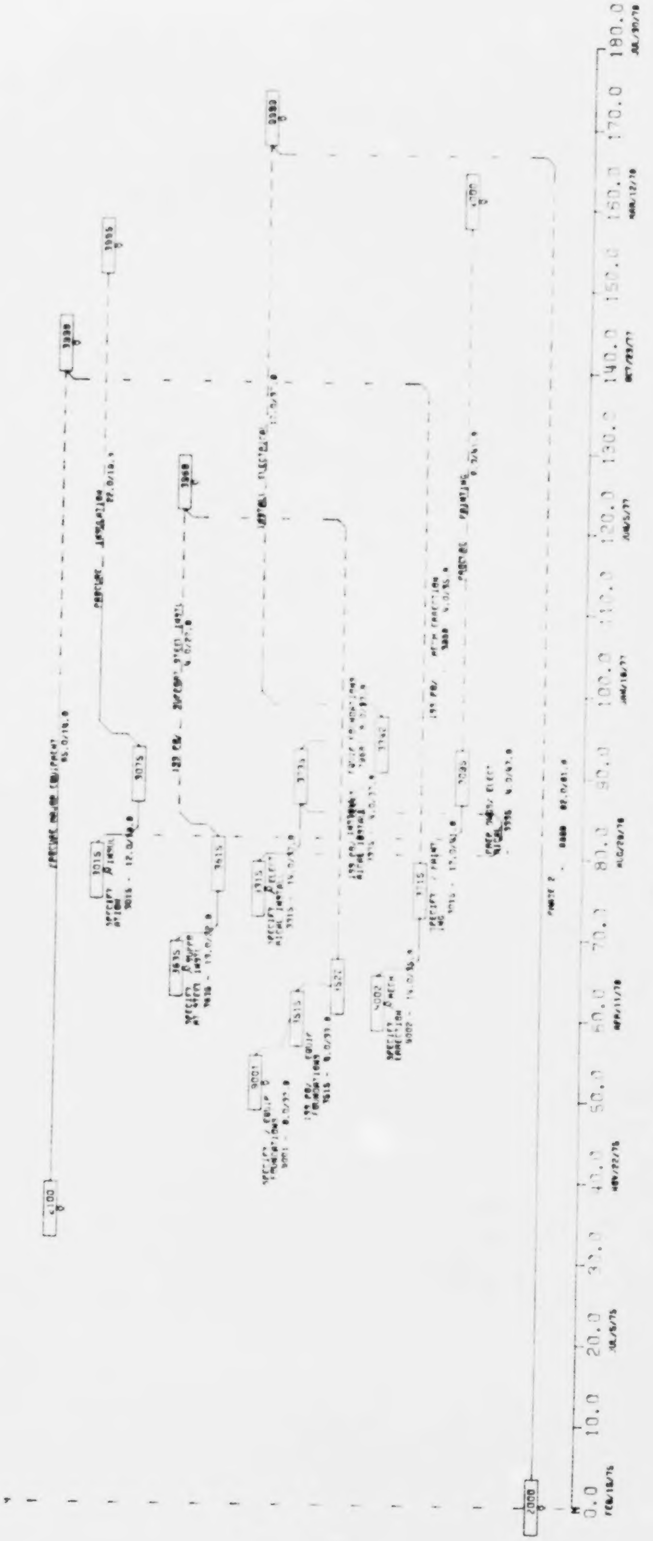


Figure 14 —Detailed design and construction schedule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40.0
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140.0
150.0
160.0
170.0
180.0
PROCURE MAJOR EQUIPMENT
PROCURE INSULATION
INSTALL ELECTRICAL
PAINTING

Figure 15 —Critical path diagram

Table 17

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FOEAS SYSTEM        ATHGAS
 REPORT 22 - ACTIVITY SCHEDULE SORTED BY  ACTIVITY SLACK                                          DATA DATE-  2/15/75

   INPUT                                      ACTIVITY                        DURATION   EXPECTED EXPECTED ACTIVITY RESP
   ORDER  PRED    SUCC                       DESCRIPTION                      EST   EFF    START   FINISH    SLACK

  1*    1  20005  2001   REC APPROVAL PHASE 2                                 4.0        15FEB75 12MAR75    -27.2 EPA
  2*    2  20005  2001   REC APPROVAL PHASE 2                                 4.0        15FEB75 12MAR75    -27.2 NEES
  3*    3  2001   2002   ISS PO/ ENGINEERING CONSULTANT                       4.0        15MAR75 09APR75    -27.2 PUR3
  4*    4  2002   2004   FINALIZE FLOW DIAGRAM                                3.0        12APR75 30APR75    -27.2 A
  5*    5  2002   2004   PREP HEAT & MATERIAL BALANCE                         3.0        12APR75 30APR75    -27.2 A
  6*    6  2004   2005   APP FLOW DIAG & HEAT BALANCE                         2.0        03MAY75 14MAY75    -27.2 ENG1
  7*   14  2005   2100   DESIGN MAJOR EQUIPMENT                              23.0        17MAY75 22OCT75    -27.2 A
  8*   19  2100   2215   SPECIFY PURCHASE RESALE EQUIP                       14.0        25OCT75 28JAN76    -27.2 A
  9*   29  2215   9001   PREP PLOT PLAN                                       2.0        31JAN76 11FEB76    -27.2 A
 10*   31  9001   9002   BUILD SCALE MODEL                                   10.0        14FEB76 21APR76    -27.2 USM
 11*   40  9002   3015   SPECIFY  / PIPING                                   16.0        24APR76 11AUG76    -27.2 A
 12*   41  3015   3998   PROCURE    PIPING                                   65.0        14AUG76 09NOV77    -27.2 PUR3
 13*    8  3998    501   PREP INSTRUCTION BOOK                               16.0        12NOV77 01MAR78    -27.2 IBW
 14*   10   501    502   PREP START-UP MANUAL                                 6.0        04MAR78 12APR78    -27.2 IBW
 15*   11   502    503   APP  START-UP MANUAL                                 2.0        15APR78 26APR78    -27.2 ENG3
 16*   13   503   9999E  SHIP START-UP MANUAL TO SITE                         4.0        29APR78 24MAY78    -27.2 CAP
 17    30  9001   9002   PREP PIPING & INSTR DIAG.                            8.0        14FEB76 07APR76    -25.2 A
 18    12   502   9999E  SHIP INSTRUCTION BOOKS TO SITE                       4.0        15APR78 10MAY78    -25.2 CAP
 19     9   501    502   APP  INSTRUCTION BOOK                                3.0        04MAR78 22MAR78    -24.2 ENG3
 20    18  3998   3999   INSTALL BOILER MODIFICATIONS                        12.0        12NOV77 01FEB78    -23.2 FS2
 21    49  3998   3999   INSTALL/ MECH EQUIPMENT                             12.0        12NOV77 01FEB78    -23.2 FS2
 22    54  3999   4000   INSTALL    INSULATION                                6.0        04FEB78 15MAR78    -23.2 FS2
 23    57  4000   9999E  INSTALL    PAINTING                                  6.0        18MAR78 26APR78    -23.2 FS2
 24    42  9002   3115   SPECIFY HIGH TEMP VALVES                            16.0        24APR76 11AUG76    -22.2 A
 25    43  3115   3998   PROCURE HIGH TEMP VALVES                            60.0        14AUG76 05OCT77    -22.2 PUR3
 26    58  9001   3635   SPECIFY  / SUPPORT STEEL                            14.0        14FEB76 19MAY76    -20.2 A
 27    59  3635   3968   PROCURE    SUPPORT STEEL                            60.0        22MAY76 13JUL77    -20.2 PUR3
 28    27  9002   3315   SPECIFY  / ELECTRICAL EQUIP                         14.0        24APR76 28JUL76    -20.2 A
 29    28  3315   3998   PROCURE    ELECTRICAL EQUIP                         60.0        31JUL76 21SEP77    -20.2 PUR3
 30    37  3968   3998   INSTALL  SUPPORT STEEL                              10.0        16JUL77 21SEP77    -20.2 FS2
 31    38  9002   3215   SPECIFY  / INSTRUMENTATION                          14.0        24APR76 28JUL76    -12.2 A
 32    39  3215   3998   PROCURE    INSTRUMENTATION                          52.0        31JUL76 27JUL77    -12.2 PUR3
 33    16  2100   2195   SPECIFY BOILER MODIFICATIONS                        20.0        25OCT75 10MAR76     -2.2 A
 34    17  2195   3998   PROCURE BOILER MODIFICATIONS                        62.0        13MAR76 18MAY77     -2.2 PUR3
 35    44  3015   3135   SPECIFY  / MISC FOUNDATIONS                         14.0        14AUG76 17NOV76      1.8 A
 36    45  3135   3042   ISS PO/    MISC FOUNDATIONS                          4.0        20NOV76 15DEC76      1.8 PUR3
 37    46  3042   3968   INSTALL/MISC FOUNDATIONS                             8.0        18DEC76 09FEB77      1.8 FS2
 38    50  3015   3055   SPECIFY  / PIPE SUPPORTS                            13.0        14AUG76 10NOV76      3.8 A
 39    51  3055   3998   PROCURE    PIPE SUPPORTS                            21.0        13NOV76 06APR77      3.8 PUR3
 40    22  2100   2294   DESIGN FUEL SYSTEM                                  14.0        25OCT75 28JAN76      5.8 A
- - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - - -
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Table 17(Continued)

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FOCAS SYSTEM        ATMGAS
 REPORT 22 - ACTIVITY SCHEDULE SORTED BY  ACTIVITY SLACK
                                                                                   DATA DATE-  2/15/75

    INPUT                           ACTIVITY                DURATION  EXPECTED EXPECTED ACTIVITY RESP
    ORDER  PRED  SUCC               DESCRIPTION             EST  EFF   START    FINISH    SLACK

 1   21   2294  3998   PROCURE FUEL SYSTEM                 60.0       31JAN76  23MAR77      5.8 PUR3
 2   20   2215  3998   PROCURE PURCHASE RESALE EQUIP       53.0       31JAN76  02FEB77     12.8 PUR3
 3   15   2100  3998   PROCURE MAJOR EQUIPMENT             65.0       25OCT75  19JAN77     14.8 PUR3
 4   52   3015  3075   SPECIFY / INSULATION                12.0       14AUG76  03NOV76     19.8 A
 5   53   3075  3999   PROCURE   INSULATION                22.0       06NOV76  06APR77     19.8 PUR3
 6   35   3635  3615   SPECIFY / SUPPORT STEEL INSTL       13.0       22MAY76  18AUG76     22.8 A
 7   36   3615  3968   ISS PO/   SUPPORT STEEL INSTL        4.0       21AUG76  15SEP76     22.8 PUR3
 8   32   9001  3515   SPECIFY / EQUIP FOUNDATIONS          8.0       14FEB76  07APR76     33.8 A
 9   33   3515  3522   ISS PO/   EQUIP FOUNDATIONS          4.0       10APR76  05MAY76     33.8 PUR3
10   34   3522  3968   INSTALL   EQUIP FOUNDATIONS          8.0       08MAY76  30JUN76     33.8 FS2
11   47   9002  3715   SPECIFY / MECH ERRECTION            14.0       24APR76  28JUL76     35.8 A
12   48   3715  3998   ISS PO/   MECH ERRECTION             4.0       31JUL76  25AUG76     35.8 PUR3
13   23   3315  3335   SPECIFY / ELECTRICAL INSTALL        14.0       31JUL76  03NOV76     37.8 A
14   24   3335  3342   ISS PO/   ELFCTRICAL INSTALL         4.0       06NOV76  01DEC76     37.8 PUR3
15   25   3342  9999E  INSTALL  ELECTRICAL                 12.0       04DEC76  23FEB77     37.8 FS2
16   55   3015  3095   SPECIFY / PAINTING                  12.0       14AUG76  03NOV76     41.8 A
17   56   3095  4000   PROCURE   PAINTING                   6.0       06NOV76  15DEC76     41.8 PUR3
18   26   3315  3335   PREP DWGS/ ELECTRICAL                4.0       31JUL76  25AUG76     47.8 A
19    7   2000S 9999E  PHASE 2                             82.0       15FEB75  08SEP76     61.8 MGT1
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unloading and conveying systems.

SWEC has assessed the cost of dismantling the existing boilers to provide space for installing the gasifier to be about equal to the recovered scrap value for the boiler tubes, shell, and steel structure within the boiler.

Note that the tabulation is based on materials take-off for instruments, switchgear, piping, and wiring. This permits the use of as low an order of accuracy as the 15 percent stated in the tabulation. Costs are tabulated on the basis of materials and equipment quotations valid for November 22, 1974, for a mechanically complete system ready for shake-down operations.

SHAKE-DOWN AND COMMISSIONING OPERATIONS

Following mechanical completion, NEES will operate the plant with the architect-engineer observing and correcting construction errors; Westinghouse and Esso will serve as consultants. Commissioning operations will include, but not be limited to

- Operating the retrofitted boiler, utilizing low-sulfur oil provided by NEES, on oil-firing to prove the reliability and turndown capability of the new dual-fuel burner system on oil.
- Operating the oil gasification system, utilizing low-sulfur oil provided by NEES, to prove the adequacy and reliability of the solids circulation, particulate removal, oil and air supply and distribution, and start-up heating systems.
- Operating the oil gasification system at design capacity and sulfur removal for at least 48 hours continuously, utilizing oil of design sulfur content (3 percent) to be provided by NEES, to confirm formally that the work has been satisfactorily completed. The architect-engineer's representatives shall be present during the performance of these tests, and upon their satisfactory completion shall so certify to the prime contractor.

- The prime contractor submitting to the architect engineer, not later than 90 days after the certification of mechanical completion, a list of additional work deemed necessary for acceptance of the construction from the architect-engineer. Upon completion of the listed work, the construction shall be considered accepted by the prime contractor.

EXPERIMENTAL TEST PROGRAM

NEES will operate the plant with Westinghouse and Esso serving as consultants. The experimental test program will include, but not be limited to, operations to investigate the following variations:

- High-sulfur oil (5 to 6 percent by weight) to be gasified at a reduced capacity to determine the sulfur removal efficiency for such feedstock
- Operating rates down to 25 percent of design capacity to be tried, to determine the practical limitations on system turndown.
- Air/fuel ratios above and below design to be attempted to determine the effects of variation from design ratio (20 percent) at full capacity and at reduced capacities.
- The regeneration reactor to be operated under sulfate generation conditions to explore the practicality of once-through operation to produce an innocuous solid waste containing all the sulfur derived from the oil feed.
- The gasifier fluidized bed to be operated at various levels to determine the effect of bed level on sulfur removal efficiency
- The gasifier to be operated with alternative temperature control methods, e.g. water/steam, to determine the effect on performance.

During the above operations, the plant will be monitored to observe the following performance parameters:

- Variation in boiler performance (steam production, boiler efficiency, boiler tube erosion/corrosion, and particulate removal effectiveness) determined during all stable periods of test operation.

- Coking in the fuel-gas system systematically monitored during all periods of stable test operation to determine, if possible, those conditions conducive to minimum coke laydown
- Pollution control effectiveness of the gasification system monitored to determine nitrogen oxide, sulfur oxide, particulate, and metal contents of the flue gases during all periods of stable test operation
- Materials of construction checked periodically to determine their effectiveness in resisting erosion from fluidized solids and corrosion from sulfides, chlorides, trace metallic elements, and so on.

Following delineation of demonstration plant operating parameters from the tests and observations described above, the gasification system will be operated as follows to provide design data for additional oil gasification installations:

- Tests will be made to determine the maximum output of fuel gas available continuously from the gasification system, using temporary additional fuel-gas combustion apparatus such as a smokeless ground flare if necessary
- An attempt will be made to operate the plant continuously for a period of 90 days at minimum fuel-gas coke laydown conditions to demonstrate reliable continuous operation for base load application during periods of heavy system demand or of large unit outage.
- Various limestones and fuel oil types will be used to determine their suitability for use in the CAFB system.
- Solids circulation and control systems will be operated to determine the most efficient parameters for pulse flow gas/solids circulation rate variations.
- The dry stone sulfation system will be operated to determine sulfur recovery and waste stone size relationships.

VII. COMMERCIAL PLANT DESIGN AND COST ESTIMATE

BASIS

The demonstration plant preliminary design and cost estimate presented in Sections V and VI of this report present the basis for the design of that 50 MW prototype facility. On the basis of certain assumptions the 50 MW design was prorated to a capacity of 200 MW. The main assumption was that a 200 MW gasifier, 12.2 m (40 ft) in diameter, was mechanically feasible and would be readily operable. Oil refinery experience, where fluid cokers and fluid cat crackers of 12.2 to 18.3 m (40 to 60 ft) diameter are operated routinely, was used as a basis for the 200 MW gasifier design. Other assumptions for the 200 MW design are that:

- All aspects of operation of the 50 MW demonstration unit will function as projected with no more than minor modifications, giving a sound basis for scale-up
- 200 MW is an appropriate commercial module size which can be built as a single train. The gasifier and absorber reactors' refractory arch distributor grids cannot be built at the approximate 12.2 m (40 ft) diameter required, so more expensive stainless steel grids must be used; these have an underside temperature limitation of 593°C (1100°F), which requires added provision for charging the gasifier with precalcined stone for every start-up following a bed removal and that such grids of this size are technically feasible.
- A regeneration/dry sulfation system with sulfur removal efficiency 90 percent, turndown ratio 4:1, same boiler efficiency, waste solids suitable for landfill disposal will be used

- There are normal air/fuel ratio, limestone makeup design rate stoichiometric, suitable limestone supply at 483 rail km (300 rail miles), waste disposal area at 48.3 truck km (30 truck miles)
- Outdoor installation is adjacent to the boiler; no change in heat conservation is allowed for
- An existing boiler using No. 6 fuel oil, with all storage and handling facilities for oil, also rail siding and car spotting equipment, is available
- Eight gas burners of double the capacity of the 50 MW plant, also eight gasifier cyclones and eight Sturtevant transfer systems for hot fines, will be used
- Real estate will be provided at no charge to the unit
- There is a clear site, with no overhead or underground obstructions; soil bearing pressure of 144 kPa (3000 psf); water table below all foundations, and seismic zone zero
- Limestone is one (1) times stoichiometric equal to 4/1.5 or approximately 2.7.

DESIGN

Except for the use of metal alloy construction in the gasifier and absorber grids, the 200 MW design closely approximates the design for the 50 MW plant as presented in Appendix E. Single unit blowers and compressors were used, and new concrete silos were included for fresh and waste stone storage. These silos were not required for the 50 MW unit because of the availability of existing coal bunkers for fresh stone storage and the use of existing ash silos for spent stone storage at the Manchester Street Station.

The 200 MW design assumes that fuel-gas line carbon burnout will be demonstrated in the 50 MW plant without the need for fuel-gas line shutoff values. It is further assumed that fuel-gas line shutoff valves will not be required for boiler safety shutdown. Fuel oil valves and gasifier steam purge piping will be used to stop the flow of fuel to the boiler and to purge with steam to eliminate the possibility of a boiler explosion. This steam purge option is based on a similar design being used by CE in the design of a 4.5 Mg/hr (5 ton/hr) coal gasification boiler fuel supply system.

COST

The estimate tabulation in this section (Table 18) is for the 200 MW size oil gasification prorated from the 50 MW estimate tabulation presented in Section VI of this report. The 200 MW costs have been prorated along conservative lines. The compressors, for example, have been estimated at the same $/W ($/hp) cost as for the one-fourth size compressors in the 50 MW estimate; and the instrumentation costs have nearly tripled for the 200 MW unit compared to the 50 MW unit. The 25 percent accuracy stated in the estimate tabulation could, because of the conservative nature of the cost proration, permit a real cost as much as 25 percent less than the $21,583,500 shown in the tabulation. This becomes especially true when the possibility of a 17 percent air/fuel ratio operation is considered, compared with the nominal 20 percent used in the 200 MW design. The lower air/fuel ratio could reduce the 200 MW bare cost to $18,300,000. Further, less conservative and possibly more realistic cost proration could reduce the $18,300,000 to $17,100,000, considering only the instrumentation and compressors. This is 15.7 percent less than the bare cost in shown in Table 18.

PERFORMANCE

Reference should be made to the discussion of performance in Section V of this report. The large 200 MW unit can expect some very nominal improvement in performance due to lower heat losses and to the slightly more efficient use of power by the larger blowers and compressors when compared to the 50 MW design. This better performance, however, is insignificant when compared with the savings in power possible with the use of a 17 percent air/fuel ratio in the 200 MW unit. An improvement

Table 18

PLANT COST SUMMARY
(200 MW Commercial Plant)[a]

Item	Material ($)	Labor ($)	Total ($)
PROCESS EQUIPMENT			
A. Towers	--	--	--
B. Boilers, S·heaters	90,000	2,500	92,500
F. Process furnaces	520,000	15,000	535,000
G. General equipment	1,800,000	110,000	1,910,000
L. Reactors	1,300,000	72,000	1,372,000
M. Drums	5,000	1,000	6,000
Q. Storage tanks	1,350,000	30,000	1,380,000
P. Pumps (including drives)	50,000	3,000	53,000
R. Compressors (including drives)	1,720,000	120,000	1,840,000
S. Stacks	--	--	--
T. Heat exchanges	150,000	5,000	155,000
TOTAL PROCESS EQUIPMENT	6,985,000	358,500	7,343,500
PROCESS MATERIALS			
C. Piping	3,300,000	1,700,000	5,000,000
D. Structures	900,000	320,000	1,220,000
E. Electrical	530,000	500,000	1,030,000
H. Buildings[b]	250,000	--	250,000
J. Civil	240,000	450,000	690,000
K. Instruments	350,000	150,000	500,000
M. Insulation[b]	300,000	--	300,000
N. Painting[b]	320,000	--	320,000
Refractory lining[b]	180,000	--	180,000
TOTAL PROCESS MATERIALS	6,370,000	3,120,000	9,490,000
TOTAL DIRECT COST (DM*+DL)	13,355,000	3,478,500	16,833,500
DISTRIBUTABLE ACCOUNTS			
V1. Insurance (excluding all risk)			2,500,000
V2. Federal and state taxes			
X. Temporary construction facilities			
Y. Field office (including insurance and taxes)			
Z. Construction tools and equipment			
O. Other distributable items			
TOTAL DISTRIBUTABLE			2,500,000
SUBTOTAL—COST OF WORK			19,333,500
U. INDIRECT ACCOUNTS			
Engineering			2,250,000
Design			
Other headquarters office			
Taxes—headquarters payroll			
Overhead allowance			
TOTAL INDIRECT (HEADQUARTERS OFFICE)			2,250,000
TOTAL, PRESENT DAY PRICES (BARE COST)			21,583,500
Fee, Escalation, and Contingency			Not included
Order of Accuracy			25%

[a] From original SWEC table.

[b] Subcontract.

in efficiency of at least 1 percent, to 2 percent, can be projected for the 200 MW unit as compared with the 50 MW unit.

OPTIONS

The 50 MW plant operation will be used to explore process options for such things as bed height, air/fuel ratio, stone size, type and rate, and so on. The 200 MW commercial plant design could be modified to:

- Use multiple train (either two, three, or four gasifiers) to provide a more flexible or a more reliable installation. A four-gasifier 200 MW design has been estimated to cost 34 percent more than a single gasifier 200 MW design. Thus, the $17.1 million bare cost discussed in the cost section above would become about $22.9 million for a four-gasifier design. A single gasifier design should have reliability comparable to that of fluid cokers in the oil industry, where one year's operation or more is common. It is possible that with sufficient experience the system reliability could approach that of fluid cat cracking, where three years between shutdowns for maintenance is expected.
- Have a larger than 12.2 m (40 ft) gasifier diameter, with a concommitant increase in capacity, such that an 18.3 m (60 ft) gasifier would provide 450 MW of fuel gas. The commercial fluid bed designs in the oil industry have gone to 18.3 m (60 ft) diameter successfully. The market survey discussed earlier in this report, however, would indicate that the market for very large single-train units would not be great. The 200 MW size or smaller would be adequate for the majority of potential installations.

VIII. DEVELOPMENT REQUIREMENTS

Development work in support of a demonstration plant program would be required in several areas in order to assure successful commercialization and to develop the full capabilities of the fluidized bed residual oil gasification process. Areas where further work is recommended include:

- The utilization of vacuum bottoms on other high metal content fuels
- The sulfur removal/spent stone processing system
- The assessment of the environmental aspects of stone disposal options
- The utilization of spent sorbent
- The demonstration of process components
- The review of safety requirements
- The development of alternative and advanced subsystems for the process.

Two subsystems are recommended for pilot-scale tests prior to completion of the demonstration plant: the pulsed flow solids transport system and the dry sulfation spent stone processing system. A full-scale cold model test of the solids transport system proposed for the demonstration plant should be carried out. This test program will provide full-scale data on the transfer system which was successfully demonstrated in the Esso pilot plant. The facility would provide design information on:

- The parameters associated with dense phase transfer of solid particles. These parameters include: particle size and size distribution, fluidization requirements (air velocity kg solids/m^3/[lb solids/ft^3 air]), mode of operation (continuous or pulsed) solids transfer rates, and distance of particle transfer.
- Transfer duct geometry including: size and shape of opening, vertical position and angle, duct material

(surface roughness - degree of corrosion) and orientation of the air nozzles

- Transfer performance with respect to variation on the operating conditions. This includes the conditions of fluidization and various conditions associated with the generation of fines (in other words, location of the fines return line, quantity of fines generated, and so forth).
- Operating control methods dependent on the solids transfer system. This includes both pressure and temperature difference induced control schemes and would involve assessing both the degree of control with respect to the dynamics of the particle transfer and an evaluation of specific equipment that could be used to effect this control.

The spent stone processing system selected for the demonstration plant should be tested in pilot-scale fluidized bed reactor units. These tests are required to check projections from laboratory tests and to develop operating experience.

Experimental work is scheduled to continue to investigate

- Stone selection and stone performance -- investigation of reactivity, attrition, trace element release, potential for high utilization for once-through operation
- Spent stone processing -- investigation of alternative processes
- Environmental impact -- investigation of spent stone analysis, leaching, and emissions; stack emissions including sulfur dioxide (SO_2), nitrogen oxide (NO_x), particulates, trace elements.

Support work recommended on the demonstration plant design includes evaluation of recommended equipment (for example, carry out pulverizer tests with spent stone from the pilot plant to obtain performance data), assessment of boiler modification options developed during the preliminary design, assessment of the cost sensitivity of critical operating and design parameters, review of safety requirements for the gasifier and gasifier-boiler interface, review of burnout and turndown procedures in light of recent pilot-plant tests carried out by Esso, and a continuing evaluation of the environmental impact and applicable regulations. Work is also recommended to develop improved understanding of the process operation:

- Continued operation of the Esso pilot plant to investigate areas such as carbon burnout procedures, control of carbon laydown, control of sorbent attrition and elutriation, temperature control schemes, improved emissions control levels, critical operating parameters (such as air/fuel ratio, bed depth, fluidization velocity, etc.), alternative limestones, operation with vacuum bottoms.
- Identification of potential alternative fuels for utilization with CAFB technology (fuel availability, market, processing techniques, economic projection, environmental impact).

In order to maintain perspective and to develop the maximum utilization of the technology, market studies must be continued. The availability of fuel and sorbent for the process must be monitored to project the impact on the market. Competitive technology must be assessed to determine its impact on the application of the oil gasification technology to the utility and industrial markets.

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LIST OF ABBREVIATIONS

A/F - air/fuel ratio
BCURA - British Coal Utilization and Research Association
$BaSO_4$ - barium sulfate
$CaCO_3$ - calcium carbonate
CAFB - chemically active fluidized bed
CaO - calcium oxide
$Ca(OH)_2$ - calcium hydroxide
CaS - calcium sulfide
$Ca(SH_3)_2$ - calcium bisulfite
$CaSO_3$ - calcium sulfite
CE - Combustion Engineering
$CH_{1/2}$ - residual hydrocarbon content of gasifier stone
CH_4 - methane
CH_3OH - methanol
CO - carbon monoxide
CO_2 - carbon dioxide
Consol - Consolidated Coal Company
COS - carbonyl sulfide
CS_2 - carbon disulfide
EPA - Environmental Protection Agency
GJ - giga joule (10^9), S.I. unit of energy
HDS - hydrodesulfurization
H_2S - hydrogen sulfide
H_2SO_4 - sulfuric acid
kg - kilogram
kPa - kilopascal, S.I. unit of pressure
kw - kilowatt
kWh - kilowatt hours
m - meter

$MgCO_3$ - magnesium carbonate
MgO - magnesium oxide
ml - milliliter
mm - millimeter
m/s - meters per second
MW - megawatt
NEES - New England Electric Systems
NO_2 - nitrogen dioxide
O_2 - oxygen
OCR - Office of Coal Research
ORD - Office of Research and Development
ORM - Office of Research and Monitoring
Pa - pascal, S.I. unit of pressure
PTG - purge transport gas
rms - root mean square
SiO_2 - silicon dioxide
SO_2 - sulfur dioxide
SO_3 - sulfur trioxide
SRE - sulfur removal efficiency
SWEC - Stone and Webster Engineering Corporation
TG - thermogravimetric